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INSTITUTE OF GAS TECHNOLOGY
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LNG: A SULFUR-FREE FUEL FOR POWER GENERATION

Final Report

Project 8926

Prepared for

NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
PUBLIC HEALTH SERVICE
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
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Introduction

I. General Purpose of the Study — Scope

The purpose of this study is to show the potential for the use of imported LNG as a low-sulfur-content fuel for power generation. To do this it was necessary to establish the availability and cost of LNG, and the cost of competitive fuels, taking into account the cost of sulfur control with high-sulfur-content fuels.

This study was limited to two seaboard locations with deepwater access. It was assumed that siting problems such as cooling water availability, aesthetic effects, land restrictions, and similar concerns will not limit placement of a single-power generating complex of 10,000-MW capacity on one site.

Currently available LNG technology — with some slight extrapolation to new and as yet untried in LNG service, marine transport technology — was used to establish costs. Competitive fuel prices were based on present levels; the cost of sulfur control was established from the ground rules set up by the National Air Pollution Control Administration.

In preparing the feasibility analysis it was assumed that all of the imported LNG was to be used solely for power generation in a conventional steam-electric power plant. No special arrangements were made nor credits taken for specialized power cycles that could utilize the cold in the LNG. The economic value of the load balancing that might be achieved if the power generating needs were integrated with a utility gas base-load supply system was not considered either.

The price of LNG at a foreign liquefaction plant is always subject to negotiation before a contract is signed; there are no "posted prices" for natural gas in foreign fields. Therefore, we have determined the cost of delivered LNG for a variety of field prices and have based the feasibility analysis on the "most likely" estimated field price.

The regulatory status of LNG importation is not yet settled. On the basis of our discussions with representatives of the various regulatory agencies that would be concerned, we have again selected the "most likely" set of regulatory circumstances as the basis for our analysis. Regulations that would have little or no effect on economics were not considered in detail.

II. Objectives of the Study

The primary objectives of this study are threefold: 1) to assess the availability of natural gas for conversion to LNG so that it could be transported to the U. S. East or West Coast, 2) to estimate the cost of transporting the available natural gas to the point of use, and 3) to estimate the cost of electric power if the imported LNG is used as power plant fuel. At the same time a detailed report on the state of the LNG art and, where possible, equipment investment and operating costs were to be prepared. The possible shifts in fuel consumption patterns engendered by imported LNG are also to be discussed.

A further objective is to assess impact of LNG as a sulfur-free fuel for power generation on the changing social, political, and economic environment.

III. Conduct of the Study

The world supply and availability of natural gas for importation into the United States has been determined. Potential supply sources were established and the cost of transporting sufficient quantities of LNG from these sources to fuel a 10,000-MW electric power generating center were estimated. The estimated cost of imported LNG was then used to calculate the bus bar cost of electric energy. This cost was compared to the cost of electric generation using the alternative fossil fuels — coal and residual fuel oil — and considering the cost of sulfur emission control. The break-even costs for imported LNG are also established.

A detailed catalog of LNG technology has been compiled from available literature sources and interviews with equipment suppliers. Investment and operating costs have been established from available literature and file data sources. Technological trends have also been assessed and the future course of technology predicted.

Finally, the possible impact of imported LNG on interfuel competition and changes in the political, social, and economic environment has been assessed.

The most recent data available were used to establish the production, consumption, and reserves of natural gas and liquefied natural gas. The principal data source for information on the U. S. natural gas industry was Gas Facts — 1967, published by the American Gas Association. Data in this publication were derived, in large part, from U. S. Bureau of Mines, Minerals Yearbook information. Data on the world natural gas industry were derived from the United Nations data, principally.

Sponsorship

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IV. Summary

Importation of LNG for Electric Power Production

A study was made of the technical and economic feasibility of importing liquefied natural gas (LNG) into the Eastern and Western Seaboards of the United States for use in electric power production. This study, carried out for the NAPCA, considered only one volume of supply — the energy needed for a 10,000-MW power center at two locations. Price and availability of gas at the remote source were based upon current information as to reserves, indigenous requirements, and marketing plans of the agencies controlling gas at the source.

LNG has only recently come to be recognized as a potentially important segment of the world energy market. It is an ideal fuel for steam-electric power generating plants because it possesses all of the favorable attributes long recognized for natural gas. It is clean burning, emitting no sulfur, no particulate matter, and, when properly controlled, relatively little nitrogen oxides or gaseous products other than carbon dioxide are discharged.

Combustion of natural gas is easily controlled, and the fuel is easily metered and handled, an important factor in keeping plant costs low. When compared to the cost of electricity generated by competitive fuels — making full allowances for the higher plant and operating costs for sulfur control with high-sulfur fossil fuels and the potentially much higher nuclear plant costs — LNG could possibly be a lower cost fuel.

Natural Gas Use and Availability

United States

The United States is by far the world's largest user of natural gas. In 1967 consumption exceeded 18 trillion CF. The U. S. S. R., the nearest competitor for gas consumption, used only about one-third of this quantity.

In the United States demand for natural gas continues to increase year by year, due in part to its nonpolluting character and consequent cost reductions resulting from its use. Research into new applications for natural gas as an energy source will probably even further stimulate our demand for this fuel.

Since 1946 the ratio of production to reserves, the so-called R/P ratio,* has steadily declined as the accumulated inventory of previously unmarketable gas was consumed. It has been projected[2] that this ratio will fall from its 1967 value of 15.8 to a value of 10.06 in 1985. Perhaps of equal significance is the capcity to produce natural gas in sufficient quantity to supply demand, including winter season peak demand. J. M. Wanenmacher[3] predicts that there will be enough total productive capacity to meet total peak demand until 1980. Beyond that time meeting the peakload must be accomplished by other means such as supplemental gas and LNG or gaseous storage.

The long-term viability of the U. S. natural gas producing industry is dependent upon the quantity of natural gas reserves ultimately available. This quantity of reserves is, in turn, dependent on 1) the gas price being sufficient to provide the incentive required to find and develop the gas, and 2) the resulting price for natural gas in the marketplace being reasonable in relation to the price of gas from other sources as well as in relation to the prices of other competing forms of energy. On this basis the Potential Gas Committee[1] estimated the remaining recoverable reserves at 976 trillion CF of gas as of December 31, 1966. IGT estimated reserves as of December 31, 1968, to be 1645 trillion CF.

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* The R/P ratio, often erroneously referred to as a life index, is, in reality, a measure of the working inventory the gas industry is willing to maintain.

The impact of extensive use of natural gas for power generation affects production, net proved reserves, necessary productive capacity, and remaining recoverable reserves. An annual consumption of approximately 740 billion CF of natural gas by a 10,000-MW power block amounts to 3.9% of the 1967 consumption. The 740 billion CF also represents 83% of the increase in consumption between 1966 and 1967. Thus, the impact of one power block in terms of current consumption would be most significant. The immediate effect on net proved reserves (0.25%) or ultimate reserves (0.045%) would be less, Table 1. Its probable effect would be to stimulate the technology of discovery, exploitation, and transportation of gas. This would result in an increase in the levels of these reserves.

World

A similar examination of world-wide natural gas use and availability in terms of the impact of its use for power generation is more difficult to make. Figures for reserves, production, and consumption are less readily available and are subject to more question. It is not possible to project production, consumption, and reserve figures based on past performance as is done for the United States.

The consumption of natural gas in the free world, exclusive of the U.S., was approximately 2.9 trillion CF in 1966 as compared to the production figure of 6.3 trillion CF, Table 2. Thus, it would seem that some five 10,000-MW power blocks could be absorbed by the free world without disturbing the current production-consumption pattern. Of course, this is not the case. The impact of gathering, liquefying, and transporting this gas is shown elsewhere in this report to be of utmost significance. The diversity of sources, quantities of reserves at each source, distance of transport, and hence cost of delivered gas preclude all but a few areas as potential supplies for United States electric power generation needs. However, the fuel for a 10,000-MW power block represents about 0.057% of world estimated proved reserves, and 0.008% of ultimate reserves.

Table 1. SUMMARY OF IMPACT OF ONE 10,000-MW POWER BLOCK

Consumption by One 10,000-MW Block = 740 billion CF/yr

World Estimated Proved Reserves – $\frac{740 \times 10^9}{1308 \times 10^{12}}$ = 0.057%

World Ultimate Reserves – $\frac{740 \times 10^9}{9200 \times 10^{12}}$ = 0.008%

United States Estimated Proved Reserves – $\frac{740 \times 10^9}{292,908}$ = 0.25%

United States Ultimate Reserves – $\frac{740 \times 10^9}{1645 \times 10^{12}}$ = 0.045%

World Consumption in 1980 – $\frac{740 \times 10^9}{68,282 \times 10^9}$ = 1.08%

U.S. Consumption in 1980 – $\frac{740 \times 10^9}{29 \times 10^{12}}$ = 2.55%

World Production in 1980 – $\frac{740 \times 10^9}{72,337 \times 10^9}$ = 1.02%

U.S. Production in 1980 – $\frac{740 \times 10^9}{26,800 \times 10^9}$ = 2.76%

World Consumption in 1966 – $\frac{740 \times 10^9}{24,654 \times 10^9}$ = 2.78%

U.S. Consumption in 1966 – $\frac{740 \times 10^9}{17,945 \times 10^9}$ = 4.12%

World Production in 1966 – $\frac{740 \times 10^9}{29,576 \times 10^9}$ = 2.50%

U.S. Production in 1966 – $\frac{740 \times 10^9}{17,491}$ = 4.23%

Table 2. WORLD PRODUCTION AND CONSUMPTION FOR 1966 AND ESTIMATED PROVEN RESERVES FOR 1967

Country	Production	Consumption	Reserves for 1967	% Reserves
	10^9 CF			
North America (exclusive of U.S.)	1,773	1,170	58,784	4.5
South America	1,464	530	51,175	3.9
Western Europe	808	840	118,219	9.0
North Africa	436	38	141,497	10.8
West Africa	105	7		
Middle East	1,483	181	297,845	22.8
Far East	165	162	33,015	2.5
Oceania	65	--	10,250	0.8
Total	6,299	2,928	710,785	54.3
U.S.	17,491	17,945	292,908	22.4
Total Free World	23,790	20,873	1,003,693	76.7
Russia and Other Communist Areas	5,786	5,781	304,563	23.3
Total World	29,576	26,654	1,308,256	100.0

Future Growth

In utilizing imported gas in the form of LNG for a 10,000-MW power block, consideration must be given not only to estimated production and consumption demands (Tables 3 and 4) of the supply country, but also to competing export projects. For Libya and Algeria, commitments already exist on future production in the form of European LNG import contracts. In South America, Brazil and Argentina may well compete for Venezuelan gas. Thus, each supply area must be evaluated in context with its own markets and competitors.

For the first six areas of potential supply, the 1975, 1980, and 1985 availabilities for LNG production/export are given in Table 5.

The areas of the world which, because of their reserves, availability of gas for export, and proximity to the U. S., are potential sources of imported gas, are —

1. Algeria
2. Libya
3. Nigeria
4. Venezuela
5. Colombia
6. Trinidad
7. Alaska
    a. North Slope
    b. Southern region
8. Canada
    a. Arctic islands
    b. Western sedimentary basin

Table 3. WORLD ESTIMATED PRODUCTION

Country	1970	1975	1980	1985
	10⁹ CF			
North America (exclusive of U. S.)	2,623	3,510	4,641	6,243
South America	1,814	2,278	2,588	3,007
Western Europe	2,372	3,806	5,200	6,418
Africa	1,351	1,970	2,629	3,461
Middle East	1,724	2,371	2,996	3,495
Far East	346	431	566	772
Oceania	337	536	696	910
Total	10,567	14,902	19,316	24,306
U. S.	20,200	23,700	26,800	29,100
Total Free World	30,767	38,602	46,116	53,406
Russia and Other Communist Areas	8,629	13,891	26,221	51,082
Total World	39,396	52,493	72,337	104,488

Table 4. WORLD ESTIMATED CONSUMPTION

Country	1970	1975	1980	1985
	10^9 CF			
North America (exclusive of U.S.)	1,761	2,342	3,030	3,920
South America	836	1,135	1,600	2,057
Western Europe	2,598	4,211	6,149	8,163
Africa	115	190	272	400
Middle East	576	846	1,155	1,408
Far East	284	442	710	1,154
Oceania	116	269	371	514
Total	6,286	9,435	13,287	17,616
U.S.	20,852	25,300	29,000	32,700
Total Free World	27,138	34,735	42,287	50,316
Russia and Other Communist Areas	8,819	14,090	25,995	51,060
Total World	35,957	48,825	68,282	101,376

Table 5. NATURAL GAS AVAILABILITY FOR LNG PRODUCTION/EXPORT

Country	1975	1980	1985
	Available gas, 10^9 CF/yr		
Algeria	336	493	697
Libya	851	997	1138
Nigeria	588	861	1219
Venezuela	296	294	300
Colombia	59	64	73
Trinidad	150	203	273
Total	2280	2912	3700

Table 5 does not include either the North Slope or Canadian production as there is no way of estimating when production will begin or what the amounts will be. The ultimate reserves of the North Slope are estimated at 200 trillion CF, while those of Western Canadian Basin and Arctic Basin of Canada are estimated at 350 and 510 trillion CF.

Table 5 shows that sufficient gas is potentially available within the presently anticipated framework to supply the needs of more than one power block. With the proper incentives, the Alaskan and Canadian sources could supply several such blocks.

Economic Feasibility of LNG for Electric Power Production

Even though it appears that gas supplies would be adequate to support several power generating blocks of the assumed size and the supply chain is technologically feasible, the real test is one of economic feasibility. Since fuel costs alone, that is, ¢/million Btu, do not tell the whole story when comparing alternative power generating concepts, all segments at bus bar electricity cost — fuel costs, operating costs, and fixed charges on plant investment — must be compared. Four different fuels were considered: coal, oil, natural gas (as distinct from imported LNG), and LNG. No direct comparisons were made with nuclear power plants as the present analysis is limited to comparison with high-sulfur fuels.

The electric generating costs were estimated for coal-, oil-, and natural-gas-burning plants. These costs include all power plant costs directly associated with electric generation and, therefore, provide a bus bar cost for each fuel considered. Transmission or distribution costs have not been considered. Overhead costs for sales or corporate expenses have not been included either except for direct supervision of power plant operation.

Investment costs were estimated for coal, oil, and gas plants. These costs were estimated for both Eastern and Western Seaboard plant sites, The investment costs therefore show a differential between coal, oil, and gas plants as well as a differential between the Eastern and Western plant locations. The Western climate allows somewhat more open construction and results in lower cost construction. These cost estimates were based upon 1970 dollars, which were then escalated for 1975, 1980, and 1985. After escalation, a constant $15/kW for desulfurizing equipment was included in the oil and coal plant estimates. These investment costs were reduced to a fixed charge per unit of energy by applying an appropriate annual fixed charge rate of 14% and considering the assumed 90% load factor.

The operating costs include two major categories: fuel costs and all other operating costs. The fuel costs component is established by multiplying the net heat rate for each type of plant by the respective fuel costs on a Btu basis. The estimated 1970 net heat rates were projected to improve at the rate of 2% for each 5-year increment into the future and were assumed to be applicable for both coastal regions. The relative cost of fuel — either natural gas, oil, or coal — was, for the purpose of this report, assumed to remain constant between the years 1970 and 1985. It was felt that the plus and minus effects, inflation vs. technological developments, would not be significantly different for any of these fuels. However, a charge of $0.10/

million Btu was added to the oil and coal fuel cost to cover the cost for desulfurizing their flue gas. The other operating costs, which include maintenance and repairs, wages and supervision, water, lubrication, and supplies, were assumed to escalate at 3%/yr.

The summations of investment charges and operating costs for gas, oil, and coal plants for plant start-ups in the years 1970, 1975, and 1985 are shown in Tables 6 and 7.

Table 6. ELECTRICAL BUS BAR COST SUMMARY FOR WESTERN SEABOARD

	Gas	Oil	Coal
		mills/kWhr	
1970			
Investment	1.51	2.22	2.58
Operating			
Fuel	3.01	3.88	3.56
Other	0.42	0.55	0.58
Bus Bar	4.94	6.65	6.72
1975			
Investment	1.74	2.53	2.93
Operating			
Fuel	2.94	3.80	3.49
Other	0.48	0.63	0.67
Bus Bar	5.16	6.96	7.09
1980			
Investment	2.00	2.88	3.34
Operating			
Fuel	2.88	3.73	3.42
Other	0.55	0.73	0.77
Bus Bar	5.43	7.34	7.53
1985			
Investment	2.31	3.29	3.82
Operating			
Fuel	2.83	3.65	3.35
Other	0.63	0.83	0.88
Bus Bar	5.77	7.77	8.05

Table 7. ELECTRICAL BUS BAR COST SUMMARY FOR EASTERN SEABOARD

	Gas	Oil	Coal
	mills/kWhr		
1970			
Investment	1.69	2.40	2.76
Operating			
Fuel	3.19	3.88	3.20
Other	0.42	0.55	0.58
Bus Bar	5.30	6.83	6.54
1975			
Investment	1.94	2.71	3.11
Operating			
Fuel	3.12	3.80	3.14
Other	0.48	0.63	0.67
Bus Bar	5.54	7.14	6.92
1980			
Investment	2.22	3.06	3.52
Operating			
Fuel	3.06	3.73	3.08
Other	0.55	0.73	0.77
Bus Bar	5.83	7.52	7.37
1985			
Investment	2.54	3.47	4.00
Operating			
Fuel	3.00	3.65	3.02
Other	0.63	0.83	0.88
Bus Bar	6.17	7.95	7.90

In all cases, natural gas, because of the current low cost and low plant cost, is the most economical fuel. However, the low fuel cost derives from the existing marketing pattern: sales of interruptible boiler fuel. If the firm gas were used, bus bar electricity costs with natural gas fuel would be higher. Tables 6 and 7 show that the next best alternative to gas-fired electric generating plants on the Eastern Seaboard is coal, but that, on the Western Seaboard, oil is a closer competitor than coal.

The cost at which imported LNG becomes an attractive economic alternative is established by determining the "break-even cost" with the preferred alternative high-sulfur fuel. This break-even cost is the cost of vaporized LNG at the boiler burner tip which will result in the same bus bar electric cost as coal on the East Coast, or oil on the West Coast. For the East Coast location the 1975 break-even cost is 49.0¢/million Btu; on the West Coast it is 51.6¢/million Btu. Table 8 provides a summary of the break-even costs.

Whether or not these break-even costs can be achieved is a function of many factors: the cost of gas at the overseas liquefaction plant; the cost of liquefaction, transportation to the United States, receiving, and storage; and finally, the cost of vaporization to a form directly comparable with natural gas supplied by pipeline.

Table 8. BREAK-EVEN VAPORIZED LNG COSTS

Year	Vaporized Cost, ¢/10^6 Btu
Eastern Seaboard	
1975	49.0
1980	51.1
1985	53.6
Western Seaboard	
1975	51.6
1980	53.2
1985	54.7

The feasibility of using LNG as a sulfur-free fuel for power generation must ultimately be based upon its delivered cost. The delivered costs of LNG from four overseas sources for the East Coast and from two for the West Coast were evaluated. This investigation considered several alternative gas purchase prices at their source, the liquefaction costs, ocean transportation costs, and storage and regasification costs at the U. S. sites. The exact quantity of overseas gas required to sustain a 741-trillion Btu/yr fuel requirement was determined. Gas purchases in the field must also include liquefaction plant fuel requirements, overseas storage boil-off, ocean transport fuel requirements and boil-off, and finally, the amount of U. S. storage boil-off and regasification fuel required. The heat content of the gas was also considered insofar as it influences the volume requirement, the plant sizes, and the number and size of tankers.

The final cost of delivered and vaporized LNG varied considerably with the alternatives evaluated. For the Eastern Seaboard supply, the vaporized cost could be as low as 40.0¢/million Btu for 5¢ Venezuelan gas, or as high as 64.1¢/million Btu for 10¢ Nigerian gas. For the Western Seaboard the range was 46.3¢/million Btu for 6¢ Canadian gas and 63.4¢/million Btu for 15¢ Alaskan gas using U. S. tankers for delivery

For both coasts supply options are available that fall below the breakeven cost of 49.0¢/million Btu for the East Coast and 51.6¢/million Btu for the West Coast. On the East Coast, Venezuelan gas, as high as 10¢/million Btu and having a delivered and vaporized cost of 46.1¢/million Btu, would be most competitive. Also for the East Coast, Algerian gas at 6¢/million Btu or Libyan gas at 8¢/million Btu, both having a delivered and vaporized cost of 48.1¢/million Btu, would be competitive.

On the West Coast, 10¢/million Btu Canadian gas, having a delivered and vaporized cost of 51.3¢/million Btu, would be competitive. Likewise, Alaskan gas at slightly less than 10¢/million Btu could also be competitive.

At 10¢/million Btu the Alaskan gas would produce a delivered and gasified cost of 52.3¢/million Btu, which would be just 0.7¢/million Btu too high, but still worthy of consideration.

Noneconomic Factors Affecting Feasibility of LNG for Electric Power Production

While any plan to import LNG for power generation must meet the test of economic viability, as several of the previously discussed possibilities do, there are a great number of other, noneconomic factors that could preclude realization of these plans. These can be subdivided into regulatory, political (both domestic and foreign), and social concerns.

Regulatory Factors

There is no doubt that imported LNG will be regulated. The principal questions are in what way and by what agencies? Were LNG to be treated merely as a new energy form that is competing solely in the liquid petroleum marketplace, a set of regulations already exists. Interstate Commerce Commission rules, Oil Import Administration regulations, Coast Guard requirements, and the whole body of controls that now affect oil importation could be revised to accommodate the slight changes LNG would necessitate. If LNG were considered to be truly natural gas, as defined in the Natural Gas Act, once again an already existing set of regulations could be revised to accommodate LNG.

However, since it does have some of the technical and market characteristics of both natural gas and liquid petroleum products, a great deal of confusion and indecision has arisen. The Federal Power Commission (FPC) suggested in its annual report for 1968, "that the Commission's regulatory role, and that of other state and Federal agencies with respect to liquefied or artificial natural gas, become the subject of legislative consideration."

Lacking the guidelines such legislation might give, one must look to the possible modes of regulation and their impact on the proposed programs.

First and foremost, it should be recognized that imported LNG is most likely to be completely interchangeable with pipeline-supplied natural gas. Therefore, it would seem most logical that the companies currently in the natural gas business, the pipeline and distribution companies, would be the first to look at LNG importation; in fact, several have. Their programs for importation would probably serve as the focal point of a program that could include the use of some of the LNG for power generation. By so doing, the gas utility could derive some load balancing benefits and the electric utility could smooth out its fuel demand curve, thus benefiting both gas and electric customers. But this type of program would immediately come under scrutiny by the utility regulatory agency, either the state Public Utilities Commission or the FPC, because utility customers would be directly affected.

A second, but less desirable, alternative would be importation of LNG by an electric utility solely for use in power generation. All economic benefits to the gas utility would be lost, and, since this would disrupt the existing gas marketing patterns, it could have a detrimental effect on the gas utility ratepayer. It is this dichotomy of view that will cause the most difficulty in resolving the utility regulation status of LNG that is imported for power generation.

Transportation of LNG will be subject to Coast Guard regulation. We expect that, by the time large-scale LNG importation becomes a reality, sufficient experience will have been gained to avoid any costly or exclusionary restrictions. Since the ICC has granted approval for interstate movements by truck, it is probable that no problems will be encountered for barge movement either.

Political Factors

Political considerations often overshadow marginal economic benefits. The major political problems this program could engender are increased dependence upon a foreign energy supply, stimulation of world trade in a commodity held in large supply by the Communist countries, displacement of employees in the existing domestic energy industry, and possibly replacement of indigenous natural gas resources by foreign supplies.

The first three will become matters of trade policy to be affected one way or the other through the imposition or lack of imposition of trade restraints such tariffs import quotas, etc.

However, competition with domestic natural gas could trigger much more immediate and directly effective response. A natural gas duty bill has been introduced in Congress to impose a duty of 3¢/1000 CF on imported natural gas. Import regulations similar to those on oil could limit the quantities imported.

Lastly, since some of the natural gas considered for import is controlled by the international oil companies, they may prefer to market this gas in areas such as Japan where they would not be competing with their own natural gas.

V. Description of LNG

When natural gas is cooled to a temperature of approximately −260°F at 1 atmosphere, it liquefies into a pure, odorless liquid — liquefied natural gas. On regasification 1 volume of this liquid is equivalent to 630 volumes of gas (630 CF of gas for every cubic foot of liquid). Liquefied natural gas is composed primarily of methane, ethane, and some of the heavier hydrocarbons. In addition, quantities of nitrogen that occur in the natural gas may also go into solution in the LNG. Because of the tremendous volume ratio advantage, the storage of gas as a liquid is economically desirable for peakshaving and transportation. This volume ratio advantage also can be usefully employed in the transportation and shipment of gas as a liquid. In its liquid form, natural gas is a colorless, odorless liquid that can be termed a cryogenic fluid. Because of its cryogenic nature, liquefied natural gas has to be stored in containers made of special materials.

There are essentially two applications of liquefied natural gas: base load and peakload. In a base-load operation, natural gas is liquefied, stored, transported to the point of use, stored there, and regasified to supply the base-load demands of the country importing the gas. The gas may be used in the gas distribution grid or for power generation. For peakload use, a gas company that needs more gas in the winter months for peakshaving purposes can liquefy and store the gas during the summer months for later usage.

Both applications require essentially the same technology of processing the gas to make it into a liquid and store it in special containers. The basic difference is that in base-load movement of natural gas, specially constructed tankers are required. Base-load plants normally have large liquefaction capacity and a large turnover of gas to justify the economics of the scheme. Peakshaving plants, on the other hand, are geared for small liquefaction capacity over a period of 200-250 days so that, by the time winter comes, there will be sufficient liquid stored in a container for use later in the winter.

There is no "standard" liquefied natural gas. Composition varies with the composition of the gas fed to the liquefaction plant and the process employed for liquefaction. Typical LNG compositions are given in Table 9 and detailed physical properties of LNG are given in Appendix B. If LNG is stored for a long period of time, its composition can change through a phenomenon called "weathering." The higher boiling components — methane and ethane — boil off preferentially, leaving behind a liquid richer in the heavier hydrocarbons. In systems where the liquid turnover is rapid, this is no problem. In other cases a small reliquefaction system on the boil-off stream maintains a constant liquid composition.

However, the lack of a standard LNG composition poses few, if any, problems to LNG usage. The output of a plant remains fairly constant; even if it did vary, the power plant burners would be able to readily accommodate the change without radical adjustment.

As a liquid fuel, LNG compares quite favorably with other liquid fuels on a Btu/lb basis. However, because of its lower density, approximately 27 lb/CF, its energy content per gal is lower than other comparable liquid fuels. Table 10 gives the heating value for several fuels.

Table 9. LNG COMPOSITIONS FROM VARIOUS COMPANIES

Company	Memphis Light Gas & Water Div.	Wisconsin Natural Gas Co.	Linde Div., Union Carbide Corp.	Tennessee Gas Transmission Co.*
Location of LNG Supply	Arlington, Tenn.	Oak Creek, Wis.	Ontario, Calif. Sacramento, Calf.	Hopkinton, Mass.
Composition, %				
CH_4	94.3	93.0	99.9+	97.34
C_2H_6	3.3	5.0	--	2.48
C_3H_8	0.5	0.5	--	0.16
C_4+	1.3		--	0.02
N_2	0.6	1.5	--	--

* Anticipated composition according to plant design calculations.

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Table 10. HEATING VALUE COMPARISON

	Heating Value	
	Btu/lb	Btu/gal
LNG (as methane)	21,200	74,200
LPG (as propane)	20,390	86,600
Gasoline	20,000	118,000
Diesel Fuel	18,700	126,500

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2. Linden, H. R., "Revised Forecast of U.S. Natural Gas Supply." Paper presented to the Natural Gas Men of Houston, Houston, December 10, 1968.
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Section 1.0. Natural Gas and LNG Resources

Natural gas reserves, production, and consumption quantities are presented for the United States and the world. The world reserve figures are based upon the latest information on net proved reserves. Net proved reserves represent that quantity of gas that should probably be produced from tested reservoirs using present technology. In addition, projections of future net reserves were made for the United States.

Historic and projected world natural gas production and consumption quantities are presented. Current and projected LNG production capacities and a projection of natural gas available for LNG export are also included.

1.1. Natural Gas and LNG Resources of the United States

In order to establish the current and projected position of natural gas in the energy economy, an analysis of the historical trends in gas supply and demand were established.

1.1.1. Natural Gas Reserves and Demands

Both demand for natural gas production and discoveries of new reserves have been increasing. Approximately 18.4 trillion CF of gas was produced in 1967 – a rise of about 5.4 trillion CF over 1960 production. There were proved recoverable reserves of 292.9 trillion CF for the same year – an increase over 1960 reserves of about 30.6 trillion CF (Figure 1.1.1.-1). This represents an annual increase in production from 1960 of about 0.75 trillion CF as against an annual increase in reserves of about 4.4. trillion CF.

However, one must recognize that a considerable portion of the natural gas produced in the United States is not marketed by the utility industry. In 1967, 202 billion therms of natural gas were produced in the United States, with only about 67% of this, or approximately 135 billion therms, actually being sold by the gas industry to the ultimate consumers. These sales include about 1.5 billion therms of manufactured, mixed, and LPG gas sales. The balance was either used in the field, sold for non-utility use such as chemical plant feedstock, consumed in the transmission and distribution system, lost, or wasted.

A more conscientious control of losses and waste has also accounted for a higher percentage in the marketed production compared with the net produced. In 1960, about 560 billion CF of gas was lost or wasted – approximately 4.2% of the total net produced – while in 1967 only 267 billion CF was lost in this manner – 1.5% of the net produced. The percent of natural gas used in the field has stayed constant at about 14%, as has the amount of gas used in underground storage. Gas lost in transmission has also declined slightly, relative to the quantity transported and taking into consideration the increase in the number of mains.

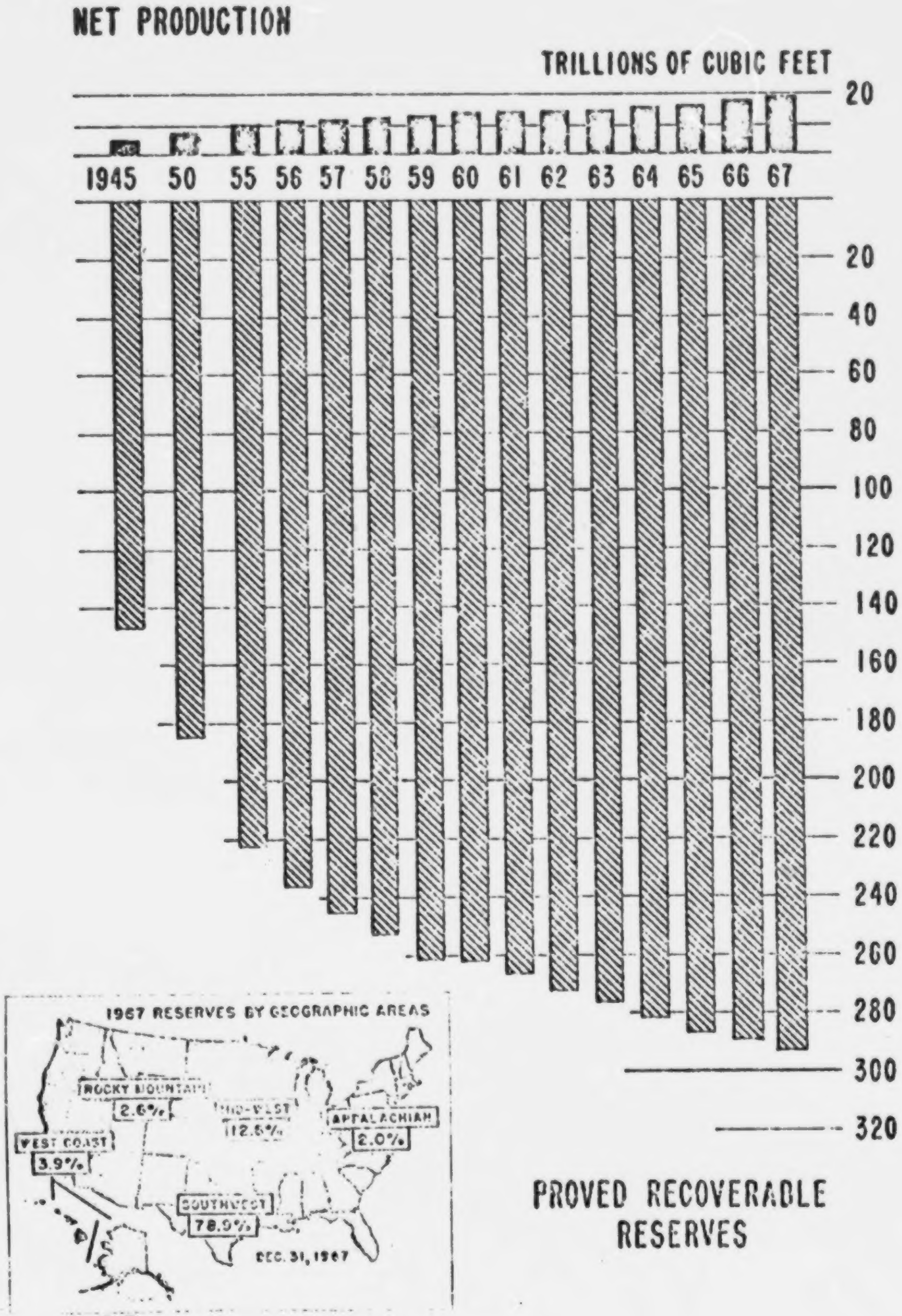

Figure 1.1.1.-1. PROVED RECOVERABLE RESERVES AND NET PRODUCTION OF NATURAL GAS IN THE UNITED STATES[1]

The United States estimated natural gas reserves were 292.908 trillion CF at the end of 1967, an increase of 3.575 trillion CF after deducting record production of 18.4 trillion CF during 1967. (See Table 1.1.1.-1.) This represented an increase of 711 billion CF in reserves over 1966. In spite of this increase, the reserves/production ratio dropped from 16.4 to 15.9. This continued decline has been a cause for concern.

Louisiana again came through with an increase big enough to push national gas reserves to another peak.

Texas had another good year, boosting its reserves by 1.8 trillion CF. This was well short of the 3 trillion CF reserve increase of the previous year, however. The most notable gains were in three areas — Dist. 8 up 2.251 trillion CF; and Dist. 7-C in the west, up 768 billion; and Dist. 4, on the Gulf Coast, up 457 billion. The Dist. 4 performance was significant in view of its position as the production leader in the state.

New field discoveries yielded 500 billion CF and new reservoir strikes 718 billion CF in this lower Gulf Coast region, exceeding all other districts and states except for Louisiana's 751 billion and 1 trillion.

On the negative side, California and Oklahoma reserves both dropped 700 billion CF due mostly to downward revisions of previous estimates. Kansas reserves also fell by 640 billion CF.

The gas reserves total included 34.2 trillion CF in the Gulf of Mexico, mainly off Louisiana.

1.1.1.1. Projected U.S. Offshore Reserves -- Atlantic Shelf

While the existence of extensive natural gas reserves offshore of Louisiana and Texas in the Gulf of Mexico has been well established and these reserves are now a major part of the U.S. energy mix, a similar, potentially as prolific source of gas lies off the northeast Atlantic Coast within easy marketing distance of potential customers for imported LNG. Similar potentials exist on the West Coast, where offshore lands underlying the Santa Barbara Channel may yield large quantities of gas and oil. These West Coast potentials will be dealt with in a later section.

Table 1.1.1.-1. ESTIMATED RESERVES FOR UNITED STATES

	Oil and Gas Journal Estimates		World Oil Estimates					IGT Estimates	
	1966	1967	1961	1964	1965	1966	1967	1966	1967
	10⁹ CF								
United States	387,000	292,596	275,000	281,251	286,469*	289,333*	292,908	289,333	292,908

* Data from table below.

State	Reserves as of 12/31/66	Changes in Reserves During 1967: Extensions	Revisions	New Field Discoveries	New Reservoir Discoveries	Net Change in Underground Storage[c]	Preliminary Net Production	Reserves as of December 31, 1967: Total Gas	Nonassocd.	Associated-Dissolved	Underground Storage[d]
	10⁶ CF (14.73 psia and 60°F										
Alaska	2,946,862	66,250	314,308	225,600	104,800	0	22,566	3,635,254	3,383,695	251,559	0
Arkansas[a]	2,599,629	267,282	17,712	4,109	46,147	2,059	125,687	2,811,251	2,528,883	261,405	20,963
California	8,474,393	205,150	(390,303)	27,700	40,997	7,265	641,405	7,723,797	3,078,508	4,455,124	190,165
San Joaquin Valley	6,250,671	146,200	(300,994)	27,700	22,000	5,103	429,497	5,721,183	2,801,684	2,822,352	97,147
Coastal Region[a]	1,269,845	2,150	(68,447)	0	3,997	2,997	121,718	1,088,824	265,362	779,759	43,703
Los Angeles Basin[a]	953,877	56,800	(20,862)	0	15,000	(835)	90,190	913,790	11,462	853,013	49,315
Colorado	1,651,406	178,743	(19,106)	75,032	1,627	768	119,207	1,769,263	1,503,440	246,754	19,069
Illinois	236,493	0	(1,334)	504	0	28,198	5,257	258,604	536	25,361	232,707
Indiana	71,207	199	510	995	0	4,079	2,209	74,781	1,830	8,394	64,557
Kansas	15,923,275	70,537	137,209	44,548	128	(3,999)	888,041	15,283,657	14,760,620	432,647	90,390
Kentucky	1,017,007	25,442	(9,312)	7,615	14,188	(7,986)	92,971	953,983	848,533	57,978	47,472
Louisiana[a]	83,684,092	2,299,890	4,400,913	750,844	1,026,915	44,761	5,917,406	86,290,009	70,899,751	15,310,629	79,629
North	5,070,784	39,074	179,449	10,647	67,935	44,761	546,160	4,866,490	3,326,732	1,460,129	79,629
South[a]	78,613,308	2,260,816	4,221,464	740,197	958,980	0	5,371,246	81,423,519	67,573,019	13,850,500	0
Michigan	773,131	3,233	(14,142)	29,489	0	1,851	32,650	760,912	90,774	87,711	582,427
Mississippi	1,668,863	39,360	50,762	1,094	4,478	(476)	167,074	1,597,007	1,268,380	322,557	6,070
Montana	620,350	123,150	8,788	99,881	3,897	13,818	32,166	837,718	550,681	138,454	148,583
Nebraska	72,757	347	(470)	61	0	573	9,476	63,792	30,762	17,083	15,947
New Mexico	14,753,142	1,045,863	187,364	52,309	10,571	(1,945)	954,839	15,092,465	11,689,327	3,387,979	15,159
Northwest	9,617,556	952,726	(14,604)	5,802	1,560	0	532,351	10,030,689	9,918,095	112,594	0
Southeast	5,135,586	93,137	201,968	46,507	9,011	(1,945)	422,488	5,061,776	1,771,232	3,275,385	15,159
New York	120,871	1,000	(3,000)	1,800	0	2,645	2,230	121,086	24,809	213	96,064
North Dakota	1,024,509	2,133	(99,198)	316	2,479	0	48,107	882,132	6,219	875,913	0
Ohio	755,215	42,250	0	6,600	0	1,166	42,500	762,731	239,989	100,006	422,736
Oklahoma	20,122,191	1,102,845	(607,499)	234,455	42,239	26,867	1,517,292	19,403,806	15,246,259	3,976,013	181,534
Pennsylvania	1,350,576	69,920	32,597	32,320	7,360	(10,637)	89,966	1,392,170	887,066	14,717	490,387
Texas[a]	123,609,326	3,745,320	2,436,673	1,542,329	1,206,492	3,506	7,128,582	125,415,064	88,507,297	36,822,517	85,250
R.R.Commn. Dist. 1	2,600,611	22,526	(75,093)	7,219	7,720	91	133,138	2,429,936	2,147,931	281,571	434
R.R.Commn. Dist. 2[a]	11,957,186	9,574	262,420	40,729	197,324	604	728,151	11,739,786	8,908,050	2,831,132	604
R.R.Commn. Dist. 3[a]	27,420,032	36,450	239,798	476,769	185,825	2,000	1,414,058	26,946,816	18,323,156	8,621,660	2,000
R.R.Commn. Dist. 4[a]	32,548,396	52,076	623,777	500,063	718,198	0	1,536,725	32,905,785	24,105,759	8,800,026	0
R.R.Commn. Dist. 5	1,487,183	48,485	(101,394)	125,902	550	(1,429)	107,316	1,451,981	1,136,052	282,073	33,856
R.R.Commn. Dist. 6	6,870,453	112,247	(3,776)	10,951	12,068	0	482,336	6,519,607	3,954,211	2,565,396	0
R.R.Commn. Dist. 7-B	889,207	29,809	(69,698)	6,343	926	(646)	108,071	747,870	249,762	462,933	35,175
R.R.Commn. Dist. 7-C	3,126,254	603,798	260,181	88,873	35,797	(152)	220,221	3,894,530	2,652,115	1,240,427	1,983
R.R.Commn. Dist. 8	17,206,882	2,480,414	424,026	186,897	26,685	0	866,641	19,458,263	12,638,629	6,819,634	0
R.R.Commn. Dist. 8-A	3,769,843	10,420	(149,506)	1,785	1,573	0	179,611	3,454,504	116,510	3,337,994	0
R.R.Commn. Dist. 9	2,768,762	8,989	140,468	2,712	2,584	2,946	230,784	2,695,677	1,892,147	792,429	11,101
R.R.Commn. Dist. 10	12,964,517	330,432	885,470	94,086	17,242	92	1,121,530	13,170,309	12,382,975	787,242	92
Utah	1,372,225	11,421	(102,185)	1,183	503	169	56,799	1,226,517	762,697	462,750	1,070
Virginia	37,586	0	4,040	0	0	0	3,828	37,798	37,798	0	0
West Virginia	2,622,237	126,481	(1,336)	20,882	11,700	11,540	211,518	2,579,986	2,174,935	58,602	346,449
Wyoming	3,594,046	111,768	237,637	9,578	130	(1,852)	265,848	3,685,459	3,202,901	455,635	26,923
Miscellaneous[b]	231,416	0	(10,050)	1,276	0	29,033	3,214	248,461	25,585	10,255	212,621
Total United States	289,332,805	9,538,584	6,570,578	3,170,520	2,524,651	151,403	18,380,838	292,907,703	221,751,275	67,780,256	3,376,172

[a] Includes offshore reserves. Total remaining recoverable natural gas reserves in the Gulf of Mexico are estimated to be 34,182,893 million CF, of which 30,192,528 million CF are nonassociated, and 3,990,365 million CF are associated-dissolved.

[b] Includes Alabama, Arizona, Florida, Iowa, Maryland, Missouri, South Dakota, Tennessee, and Washington

[c] The net difference between gas stored in and gas withdrawn from underground storage reservoirs, inclusive of adjustments and native gas transferred from other reserve categories

[d] Gas held in underground reservoirs (including native and net injected gas) for storage purposes.

( ) Denotes negative volume.

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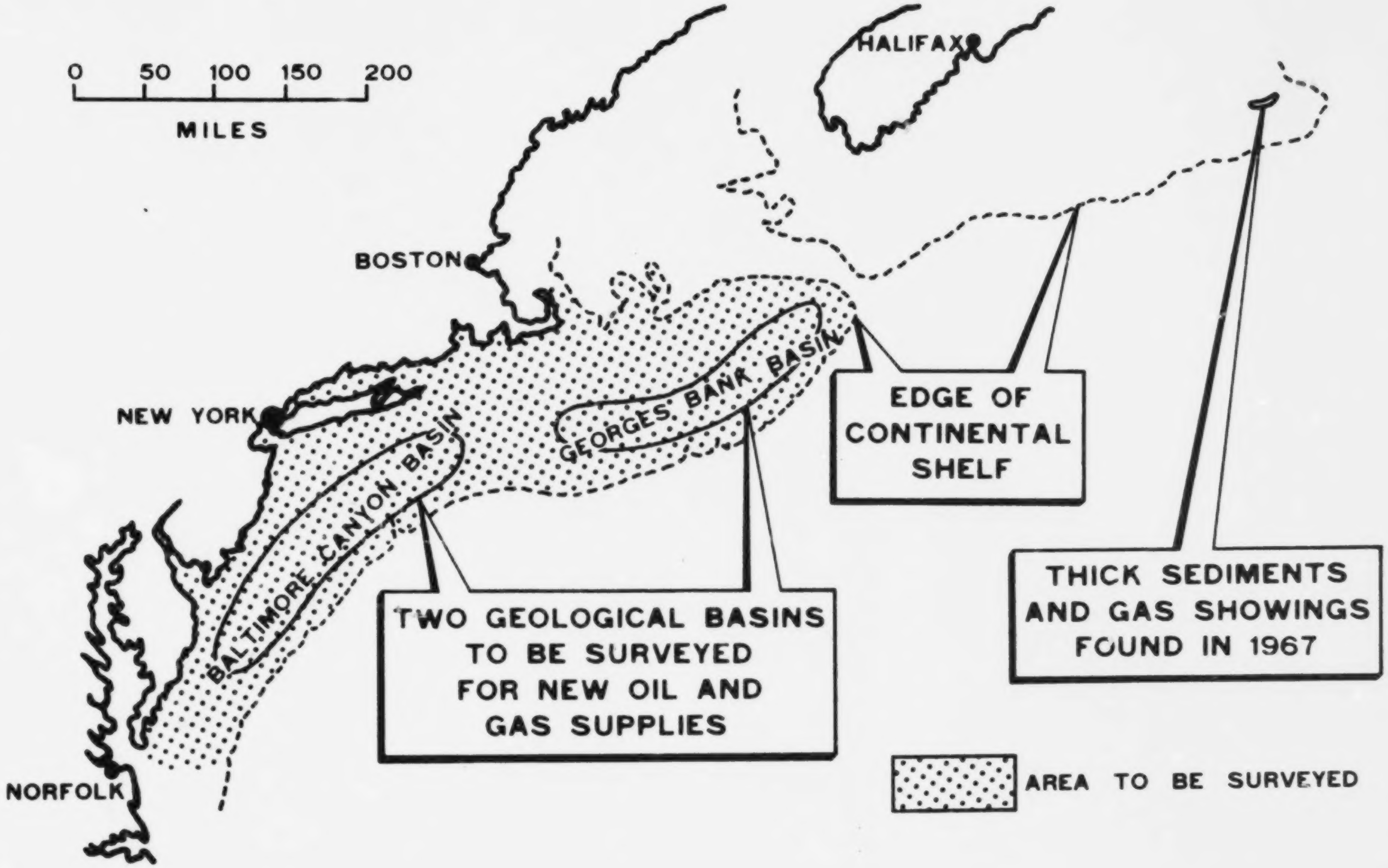
0
50
100
150
200
MILES
HALIFAX
BOSTON
NEW YORK
NORFOLK
GEORGES BANK BASIN
BALTIMORE CANYON BASIN
EDGE OF CONTINENTAL SHELF
THICK SEDIMENTS AND GAS SHOWINGS FOUND IN 1967
TWO GEOLOGICAL BASINS TO BE SURVEYED FOR NEW OIL AND GAS SUPPLIES
AREA TO BE SURVEYED

Figure 1.1.1.-2. GRAVITATIONAL AND MAGNETIC SURVEY AREAS[3]

With the rapid evolution of deepwater drilling and pipeline technology, increasing interest is being paid to the Continental Shelf regions. Conventionally, the Continental Shelf is defined as the submerged part of a continent out to a water depth of 600 feet. In the North Atlantic Continental Shelf region, the Continental Shelf averages approximately 100 miles wide for a distance extending 750 miles from Norfolk, Va., to the outer tip of the Georges Bank due east of Boston. This area, totaling 75,000 square miles of sedimentary rock, is larger in area than the Continental Shelf off Louisiana and Texas.

In spite of its apparent attractiveness, this region has been the subject of only a minor exploratory effort. The geophysical exploration that has been carried out has been principally limited to aeromagnetic surveys.

A gravitational and magnetic survey of the area is now under way. The survey, under the sponsorship of a number of major utilities and oil and gas companies, is expected to be completed by the end of 1968. The survey is not a detailed geological survey; it is a determination of the gravitational and magnetic variations which will indicate the features of the undersea formations. This will show where and what kind of rock formations exist, where the basins are, and the existence of what might be potential oil- and gas-bearing structures. The project encompasses 16,000 nautical miles of tracklines in two segments — the Georges Bank and Baltimore Canyon (Figure 1.1.1. -2).

Development of the potentially prolific off-shore areas may be slowed because of increasing concern over control of drilling and well completion techniques. Blowouts and other accidents resulting in release of oil contamination may lead to stringent restrictions which could increase the cost of off-shore drilling and production.

1.1.1 1.1.1.2. Current Natural Gas Demands

In 1966[1] the United States economy consumed a total of about 56 quadrillion Btu of energy. Of this, dry natural gas contributed about 20 quadrillion Btu, some 36%, making it second only to petroleum as an energy source (Figure 1.1.1.-3).

1.1.1.2.1. Projected Production and Consumption

The following table summarizes the projected U.S. natural gas production[4] and consumption through 1985.

Table 1.1.1.-2. PROJECTED PRODUCTION AND CONSUMPTION FOR THE UNITED STATES

Production				Consumption			
1970	1975	1980	1985	1970	1975	1980	1985
10^9 CF							
20,200	23,700	26,800	29,100	20,852	25,300	29,000	32,700

1.1.1.3. Natural Gas Supplemental Requirements

From the forecast data on production and consumption it is possible to compute supplemental gas requirements. One method assumes that the working inventory of natural gas reserves cannot be allowed to fall below a level at which investors and consumers would lose confidence in the gas industry's capability to meet market demands at competitive prices for the foreseeable future.

If reserves / production (R/P) ratios of 12 and 10 years are utilized, the amount of supplemental gas required to maintain a constant R/P ratio of the specified value may then be calculated. The results are shown in Table 1.1.1.-3.

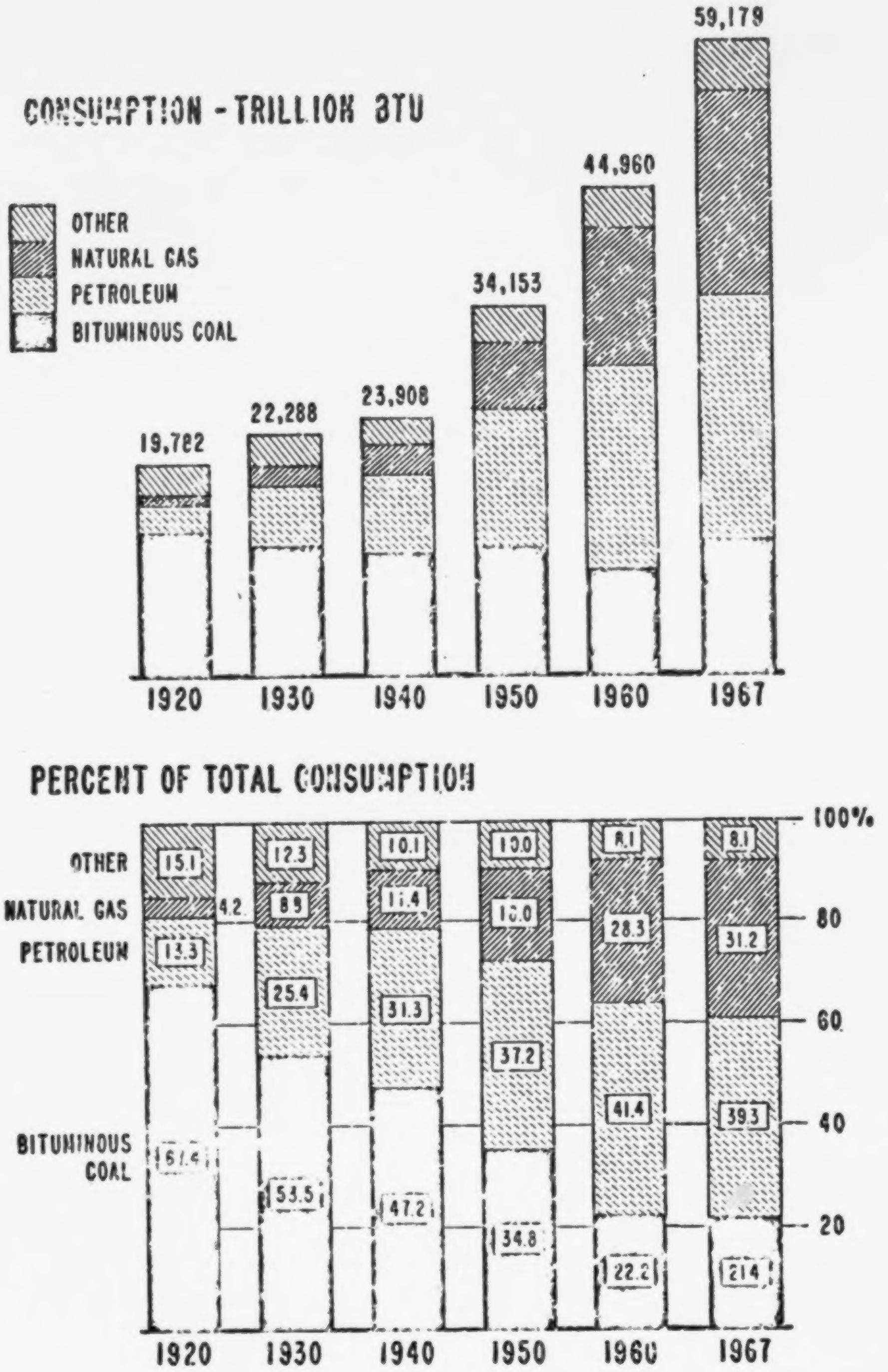

Figure 1.1.1.-3. UNITED STATES ENERGY CONSUMPTION[1]

Table 1.1.1.-3. SUPPLEMENTAL GAS REQUIREMENTS

End of Year	Natural Gas Reserves Required to Maintain R/P of		Supplemental Gas Requirements Required to Maintain R/P of	
	12	10	12	10
	10^{12} CF		10^{12} CF/yr	
1970	303.33*	303.33*	0	0
1975	311.68*	311.68*	0	0
1980	321.60	308.12*	1.12	0
1985	349.20	292.57*	4.72	0

* Computed values of R/P higher than minimum required for desired R/P value without need for supplemental gas.

1.1.2. U.S. LNG Resources

1.1.2.1. Current LNG Production

Present U.S. LNG production comes from the following plants given in Table 1.1.2.-1.

Table 1.1.2.-1. U.S. PLANTS CURRENTLY PRODUCING LNG

Plant Location	Liquefaction Capacity, 10^6 CF/day	Use
1. Wisconsin Natural Gas Co., Oak Creek, Wis.	1.0	Peakshaving
2. San Diego Gas & Electric Co., Chula Vista, Calif.	2.0	Peakshaving, remote supply
3. Alabama Gas Corp., Birmingham, Ala.	4.65	Peakshaving
4. Transcontinental Gas Pipe Line Co., Hackensack, N.J.	7.0	Peakshaving
5. Tennessee Gas Pipeline Co., Hopkinton, Mass.	18.0	Peakshaving
6. City of Memphis, Tenn.	5.0	Peakshaving
7. Brooklyn Union Gas Co., New York,	5.7	Peakshaving
8. Northwest Natural Gas Co., Portland, Oreg.	2.5	Peakshaving

1.1.2.2. Projected LNG Production

Figure 1.1.2.-1 presents the U.S. historic as well as projected LNG liquefaction capacity. The projection is based on an annual increase of capacity of 10% from 1970 through 1985. Growth rates are predicated on growing demands for peakshaving LNG, only. Other uses as fuel supply to remote areas, vehicular fuel, etc., may significantly accelerate this growth rate.

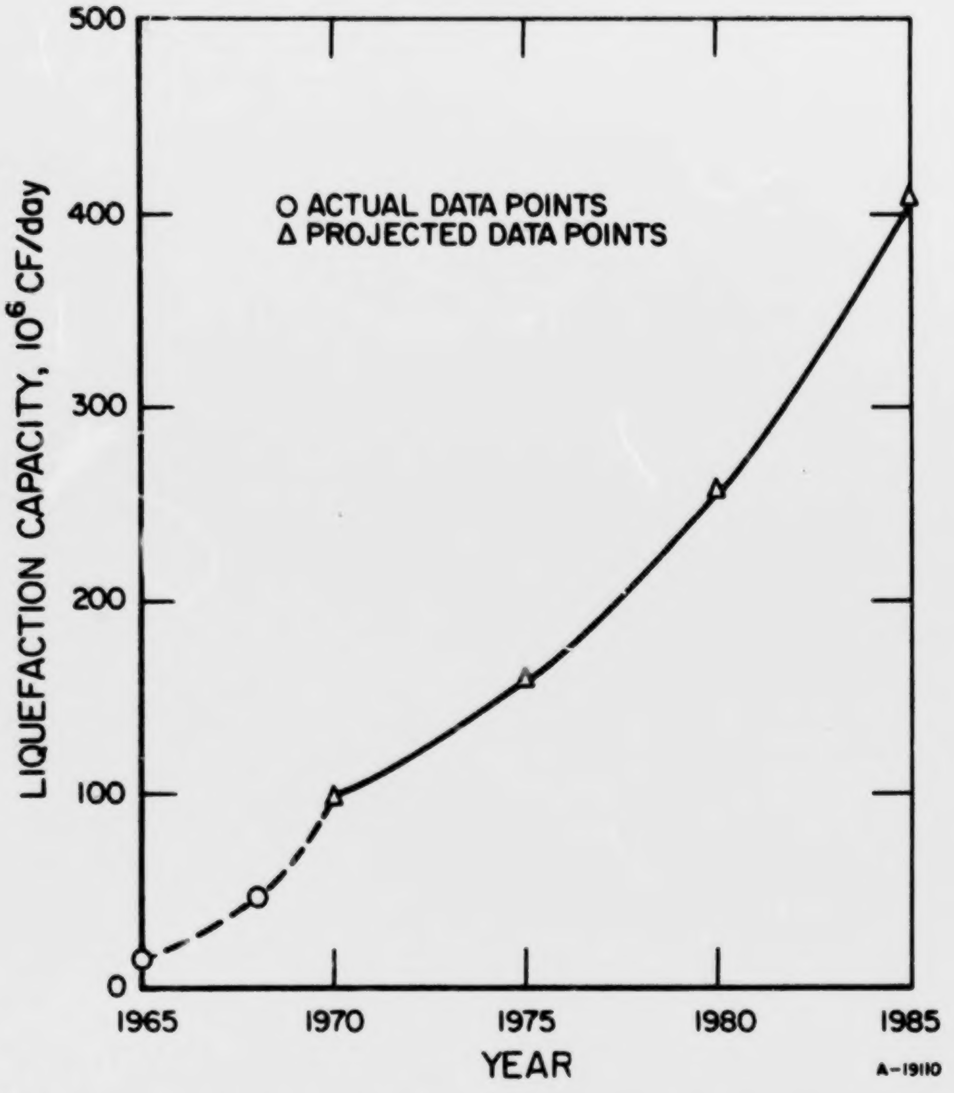

Figure 1.1.2.-1. PROJECTED GROWTH OF U.S. NATURAL GAS LIQUEFACTION FACILITIES

1. 2. World Natural Gas Reserves

1. 2. 1. Current World Gas Reserves

1. 2. 1. 1. North and Central America (Excluding the U. S.)

1. 2. 1. 1. 1. Canada

In 1967, the reserves of marketable natural gas in Canada showed an increase of 2.2 trillion CF. (See Table 1.2.1.-1.) The area with by far the largest increase in reserves was Alberta, which was followed by British Columbia, Saskatchewan, and Ontario.

Although operators proved more new reserves of natural gas than the production for the year, thereby making considerable additions to reserves, the additions were not as large as those in 1966 when large reserves were found in the Rainbow-Zama area and other districts of Alberta. The largest increases of 1967 were found in the following fields: Boqzeau Rover Elkton, up 310 billion CF; Kaybob, up 260 billion CF; Edson Elkton, up 220 billion CF; and Madison Hot Sandstone, up 200 billion CF. In addition to the Alberta additions, 14 gas discovery wells were drilled in Ontario. However, the volume of reserves added by these wells was small.

Three governmental actions that will affect future exploration activities are 1) the resolution of jurisdiction over offshore mineral rights, 2) the exportation of natural gas to the U.S., and 3) the formation of a consortium of the government and 20 private companies for an exploration program in the Arctic islands.

The long simmering question of jurisdiction over offshore mineral rights was partially resolved last year when the Supreme Court of Canada handed down a decision to the effect that mineral rights offshore British Columbia were in fact the property of the Federal Government. Needless to say, the government of the province of British Columbia is contesting this decision, and it seems likely that some compromise solution will be sought whereby the Federal and Provincial Governments will share these privileges. The eastern Canadian offshore mineral rights have not been resolved. Authorities in Canada feel that up to four points of view must be considered. In addition to the Provincial and Federal view points, are those of England and France, who may choose to intervene.

Another major Federal decision emerged when the National Energy Board (NEB) of Canada declined to permit the Westcoast Transmission Co.,

Table 1.2.1.-1. ESTIMATED RESERVES FOR NORTH AMERICA (Excluding the U.S.)

	Oil and Gas Journal Estimates			World Oil Estimates				IGT Estimates	
	1966	1967	1961	1964	1965	1966	1967	1966	1967
Country	10^9 CF								
Canada	45,000	48,000	36,000	43,391	44,372	43,450*†	45,682*†	43,450*	45,682*
Mexico	13,500	11,596	10,000	11,510	11,250	11,723	11,851	11,400	11,600
Trinidad	1,000	1,211	1,200	1,200	1,200	1,300	1,502	1,300	1,502

* Estimated on approved marketing basis.

† Data from the table below.[2]

	Proved Remaining Marketable at Dec. 31, 1966	Changes Due to					Proved Remaining Marketable at December 31, 1966						
		Revisions	Extensions	1967 Discoveries*	Net change in Underground Storage	1967 Net Production†	Total	Nonassocd.	Associated	Dissolved	Underground Storage	Change During Year	Probable Remaining Marketable at Dec. 31, 1967‡
		10^6 CF (14.65 psia and 60°F)											
Northwest Territories	117,320	-9,622					107,698	107,698				-9,622	298,040
British Columbia	7,265,690	318,121	84,499	286,652		202,217	7,752,745	7,473,745	140,464	138,579		487,055	8,871,241
Alberta	35,135,103	1,224,438	1,178,643	312,707	-1,052	959,408	36,890,431	29,018,932	4,652,950	3,200,549	18,000	1,755,328	40,680,167
Saskatchewan	729,278	35,037	4,860		1,038	41,246	728,967	349,848	306,076	62,932	10,111	-311	899,599
Manitoba													
Ontario	200,725	4,026		4,100	4,134	13,000	199,985	101,088			98,897	-740	202,897
Other Eastern Canada	1,979	424				178	2,225	1,425		800		246	2,225
Total Natural Gas	43,450,095	1,572,424	1,268,002	603,459	4,120	1,216,049	45,682,051	37,052,693	5,099,490	3,402,860	127,008	2,231,956	50,954,169

* Limited areas are generally assigned to new discoveries even though much larger areas may eventually be classified as proved.

† The 1967 net production figures were compiled from records of actual production for whatever period they were available for, with estimates for the remainder of the year.

‡ Includes proved reserves.

Ltd., to export natural gas to the U.S. under a contract price set by the Federal Power Commission (U.S.).

The price issue stems from a 1954 contract under which Westcoast began selling gas to El Paso Natural Gas Co. at 22¢/1000 CF. Last year Westcoast and El Paso negotiated a new contract that would have increased the flow of gas to 500 million CF/day at about 32¢/1000 CF, yielding an average of 27¢/1000 CF for the entire 500 million.

The NEB approved this as a fair and reasonable price, but the FPC turned it down, ruling that the 1954 contract should continue and that El Paso could pay no more than about 29-1/2¢ for the additional amount. In declining the Westcoast application, the NEB based its ruling on the view that the U.S. agency was trying to extend its internal pricing policies to Canada, which was unacceptable.

Westcoast has applied for and obtained an extension of authority to export an initial 100 million CF/day for 6 months. El Paso in turn has offered to "borrow" the gas for repayment later.

Much more is at stake, however, than the difference in prices between the two regulatory bodies: There is the fundamental difference regarding principles and politics, differences that could have a major effect in determining the extent of future Canadian gas exports to the western U.S. Affected by this decision will be the application of Union Gas Co. of Canada, Ltd., to import sizeable volumes of natural gas from Panhandle Eastern Pipe Line Co. into southwestern Ontario. The price of this gas is approximately 20% more than that paid by U.S. companies in neighboring Michigan.

The Canadian Government has subscribed 45% of the initial $20 million to be raised by Panarctic Oils Ltd. for a large exploration program across 44 million acres of Canada's Arctic islands. A group of 20 private oil and mining companies will provide the balance of the financing.

The Arctic islands and offshore areas of the Maritime Provinces and British Columbia may well hold the majority of Canada's potential reserves.

Shell Oil Co. in 1967 drilled the first two offshore wells off the west coast of Canada; both were dry and abandoned. One reached a depth of over

12,000 ft. Shell holds 13 million offshore acres, and will drill a total of 10 wells before the end of 1969.

In eastern Canada, because of the continued interest in offshore prospects of the Maritime Provinces, offshore holdings increased. In 1967 Mobil Oil Canada, Ltd., completed its deep test on Sable Island, on the Continental Shelf about 180 miles southeast of Halifax, Nova Scotia. It was the first offshore exploratory well drilled on the Nova Scotia shelf. Drilled to 15,106 feet, the well was reported to have had a show of gas, but it was abandoned as a dry hole.

1.2.1.1.2. Mexico

The exploratory activity of Mexico were mainly concentrated along the coastal plain of the Gulf of Mexico, where the four known oil and gas provinces are located. Within the coastal plain and offshore areas, the intensified exploratory efforts have brought satisfactory results. Although a slight decrease in reserves is indicated (Table 1.2.1.-1), an increase is expected after evaluation studies are completed and all figures are known.

1.2.1.1.3. Trinidad

No significant activity in gas exploration was reported from Trinidad.

1.2.1.2. South America

Presented below are estimates of the proved reserves for South America. Since the period 1966-6 includes two sources of data, an IGT estimate is made which weigh these values plus additional data from other sources.

The year 1967 marked a leveling off of the increases in reserves for Argentina and Peru. The steady decline in Venezuelan reserves is representative of the area's need to provide incentives for oil and gas exploration. In exploration, oil is still the main target, with gas prospects generally being secondary ones.

On a more positive note, there are indications that the area may be entering a period of increased exploration and development activity. The

Table 1.2.1.-2. ESTIMATED RESERVES FOR SOUTH AMERICA

	O&G Journal Estimate		World Oil Estimate					IGT Estimate	
	1966[2]	1967[1]	1961[7]	1964[6]	1965[5]	1966[4]	1967[3]	1966	1967
Country	10^9 CF								
Argentina	8,000	7,960	6,000	6,500	8,000	9,000	8,000	8,800	8,000
Bolivia	5,000	4,050	250	2,000	2,000	2,000	3,000	2,000	3,200
Brazil	500	877	350	500	671	950	1,000	955	1,000
Chile	2,000	3,530	1,800	3,500	2,800	3,000	3,000	3,000	3,000
Colombia	3,500	4,030	1,400	4,000	4,500	4,000	4,000	4,500	4,000
Ecuador	50	65	75	65	65	75	100	75	100
Peru	1,000	2,012	725	2,000	2,000	2,500	2,000	2,500	2,000
Trinidad	1,000	1,211	1,200	1,200	1,200	1,300	1,502	1,300	1,375
Venezuela	30,000	31,770	33,000	30,745	30.019	29,177	27,833	30,000	28,500

[1] Oil and Gas Journal 66, 119 (1967) December 25.

[2] Oil and Gas Journal 64, 111 (1966) December 26.

[3] World Oil 165, 47 (1968) August 15.

[4] ibid.

[5] World Oil 163, 95 (1966) August 15.

[6] World Oil 162, 91 (1965) August 15.

[7] World Oil 165, 47 (1968) August 15.

known areas for possible further investigations are extensive.

Table 1.2.1.-2 gives the estimated reserves for South America.

1.2.1.3. Western Europe

Both on and offshore, increases in natural gas reserves are continuing to justify the interest of petroleum operators. In contrast to the slowly declining oil production, natural gas production rose rapidly. The slow down in drilling in 1967 was due to pending legislation in the Netherlands, Sweden, Italy-Sicily, and Portugal. (See Table 1.2.1.-3.) This legislation, dealing primarily with the opening of offshore areas, has affected the reserve picture of Western Europe, especially in Italy where, within 1 year, 2.12 trillion CF of gas reserves were discovered in the upper Adriatic Sea. In spite of the Adriatic Sea strikes and those in the Aquitaine basin of southern France, the North Sea remains the area of prime interest. Even in the face of abandonment of drilling in the German sector after 11 unsuccessful wildcats, the overall basin picture remains bright, with new strikes in the United Kingdom and Norwegian sectors.

Table 1.2.1.-3. ESTIMATED NATURAL GAS RESERVES FOR WESTERN EUROPE

Country	Oil and Gas Journal Estimates		World Oil Estimates					IGT Estimates	
	1966	1967	1961	1964	1965	1966	1967	1966	1967
	10⁹ CF								
Austria	960	965	850	936	1,000	1,200	1,000	935	940
France	7,000	8,500	9000	9,000	8,800	8,600	10,500	3,598	5,300
West Germany	7,800	8,500	1500	6,000	7,000	8,500	8,500	7,880	7,100
Italy	4,422	5,500	4650	4,800	4,700	5,000	5,000	4,985	7,119
The Netherlands	57,500	82,000	2600	43,000	45,000	63,000	68,000	65,300	71,000
Sicily	(Included under Italy)		350	350	350	350	350	350	350
Spain			--	--	5	5	10	5	10
United Kingdom	10,000	25,000	5	7	10	10,000	25,000	10,000	25,000

1.2.1.3.1. Austria

In 1967 gas production in Austria showed a slight decline and no major gas reserves were discovered. This somewhat dismal picture was matched by West Germany. The abandonment of the North Sea area by Conoco and the rather minor onshore additions to reserves, coupled with a reduction in inland crude oil production incentives, does not provide much encouragement for future additions to reserves.

1.2.1.3.2. France

In France, the greatest activity was in the Aquitaine basin. This basin, lying to the north of the Pyrenees, contains the large Laca field. Since the currently estimated reserves of 2 trillion CF for the new strikes are conservative and will be revised upward as drilling proceeds, it is not unlikely that reserves in the Aquitaine basin will exceed those of the Laca deep.

1.2.1.3.3. Italy

The main exploration target of Italy is gas rather than oil. All recent energy forecasts show a rapid growth of gas use. The main focal point of this exploration will be the promising potential gas-bearing structures of northern Adriatic waters.

1.2.1.3.4. The Netherlands, Norway, and United Kingdom

For the Netherlands, Norway and the U.K., the North Sea and near-shore areas remain the most potential source of new reserves.

1.2.1.4. North Africa

Even though the Middle East War affected petroleum production in 1967, exploration activity increased 27.5% over 1966. (See Table 1.2.1.-4.) During 1967, wildcat completions decreased numerically, but the success percentage increased to 22.6%, including three major gas fields. The recent activities of North African governments in the petroleum area have reflected their desires for a more direct role in exploration and production. In the future, this trend should continue.

Table 1.2.1.-4. ESTIMATED RESERVES FOR NORTH AFRICA

Country	Oil and Gas Journal Estimates 1966	Oil and Gas Journal Estimates 1967	World Oil Estimates 1961	World Oil Estimates 1964	World Oil Estimates 1965	World Oil Estimates 1966	World Oil Estimates 1967	IGT Estimates 1966	IGT Estimates 1967
	10^9 CF								
Algeria	140,000	140,000	50,000	60,000	65,000	80,000	100,000	105,000	120,000
Egypt	530	1,300	425	500	500	750	1,000	750	1,500
Libya	7,000	15,000	3,700	5,000	7,500	8,000	11,000	12,400	14,000
Morocco	25	23	10	17	17	18	18	20	20
Tunisia	500	750	10	10	300	400	500	400	500

1.2.1.4.1. Algeria

As an example of government activity, Algeria announced its intention to nationalize the petroleum industry on May 14, 1968. Société Nationale de Transport et de Commercialization des Hydrocarbons (Sonatrach) was formed to develop and exploit the hydrocarbon resources of Algeria.

Although one of the promising new gas fields, at Gassi el-Aden, has reserves estimated at 1.06 trillion CF, the main obstacle to Algerian gas development is price. As an example, Algeria gas delivered in the United Kingdom costs about $0.65/1000 CF (as LNG), whereas North Sea gas will cost $0.278/1000 CF. However, Algeria will continue to be one of the major sources of natural gas exported as LNG.

1.2.1.4.2. Egypt

Egypt, which suffered great losses in its Sinai Peninsula fields during the Middle East War, has recently announced the discovery of a major field at Aru Madi in the Nile Delta. Gas from this field, which has estimated reserves from 0.35 to 2.5 trillion CF, will be piped to both the northern area of the Delta and to the southern industrial area of Cairo and the Helwan. Increases in consumption are expected to keep pace with increases in reserves.

1.2.1.4.3. Libya

The continued growth of Libyan petroleum industry is attributed to its favorable geography and political stability as well as its extensive reserves. At present, most of the gas, which is produced as associated gas with oil, is flared. With the completion of the Esso LNG plant at Marsa El Brega, 345 million CF/day of gas will be liquefied for export.

1.2.1.4.4. Morocco and Tunisia

Little activity of note has occurred in 1967 in either Morocco or Tunisia. The main emphasis in these two countries has been on development of existing fields.

1.2.1.5. West Africa

Petroleum activity continued at a high level in West Africa until mid-1967 when civil war broke out in Nigeria. Even with the difficulties imposed by the war, exploration was at the same level as 1966. (See Table 1.2.1.-5.)

In Cabinda and Angola exploration activity increased sharply. In an offshore area 40 miles north of the Congo River in Cabinda, 13 successful wells were drilled.

After many years of relative disappointment, the large new fields of Gamba and Anguille have come into their own. The associated gas for these fields accounts for the 50% increase in gas production in 1967.

In both the Congo and Senegal, little exploration activity is taking place.

In Eastern Africa, several gas strikes were made in Mozambique. In addition to the potentially promising areas on- and offshore in Mozambique, Madagascar is also thought to possess good possibilities. (See Table 1.2.1.-5.)

1.2.1.6. Middle East

In 1967 one new gas field was discovered in the Rubal Khali area of Saudi Arabia. Other than this, the new additions were in associated gas. Offshore activity in the Persian Gulf area has increased over the past few years resulting in the discovery of major fields. These include Iran's Sassan and Cyrus fields, Abu Dhabi's Zakum field, and Daubai's Fatch field. Onshore discoveries were made in Saudi Arabia and in the Neutral Zone. (See Table 1.2.1.-6.)

Table 1.21.-5. ESTIMATED RESERVES FOR WEST AND EAST AFRICA

Country	Oil and Gas Journal Estimates 1966	Oil and Gas Journal Estimates 1967	World Oil Estimates 1961	World Oil Estimates 1964	World Oil Estimates 1965	World Oil Estimates 1966	World Oil Estimates 1967	IGT Estimates 1966	IGT Estimates 1967
	10^9 CF								
West Africa									
Angola	150	150	7	120	100	100	25	20	25
Cabinda	(Included under Angola)		--	--	--	--	150	80	150
Congo (Brazzaville)	--	--	5	2	2	2	2	2	2
Gabon	5000*	5000*	30	150	160	200	300	200	300
Nigeria	4000	4000	250	1700	3000	4500	3000	4500	4000
Senegal	--	--	6	--	--	--	--	--	--
East Africa									
Mozambique	1000	1000	--	--	--	--	--	1000	1000

* May be due to estimate of associated gas for Gamba and Anguille fields.

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Table 1.2.1.-6. 1967 ESTIMATED RESERVES FOR MIDDLE EAST

Country	Oil and Gas Journal Estimates	World Oil Estimates	IGT Estimates
		10^9 CF	
Abu Dhabi	8,000	6,000	15,000
Bahrain	100	200	200
Dubai	--	700	700
Iran	109,500	100,000	214,000
Iraq	20,000	21,000	21,000
Israel	70	37	45
Kuwait	34,500	35,000	35,000
Neutral Zone	3,500	6,000	6,000
Oman	1,500	1,500	1,500
Qatar	7,500	7,000	7,500
Saudi Arabia	35,000	50,000	50,000
Syria	500	500	500
Turkey	500	300	400

1.2.1.7. Far East

Presented below is a summary of the estimated proved natural gas reserves for the Far East, as of the end of each year, for 1961, 1964, 1965, 1966, and 1967. Far Eastern industrial development is being based primarily on gas reserves, especially in Pakistan and Afghanistan. Japan is rapidly developing several small fields. Both East and West Pakistan are actively constructing new gas pipelines.

Table 1.2. 1. -7. ESTIMATED RESERVES FOR FAR EAST[5]

Country	1961	1964	1965	1966	1967
	10^9 CF				
Afghanistan	--	2,000	3,000	4,000	5,000
Burma	150	150	200	200	200
China	2,000	3,000	3,500	3,500	3,500
India	745	1,500	2,000	2,000	2,065
Japan	500	700	700	800	800
Pakistan	15,000	18,750	20,000	20,300	20,500
Taiwan	10	100	200	964	950
Thailand	1	--	--	--	--

1.2.1.8. Oceania

Tabulated below are the estimated proved natural gas reserves for Oceania, as of the end of each year, for 1961, 1964, 1965, 1966, and 1967. Oceania's principal gas reserves are in Australia, where pipelines are presently under construction. New Zealand is also constructing a major pipeline scheduled to be in operation about the middle of 1969.

Table 1. 2. 1. -8 ESTIMATED RESERVES FOR OCEANIA[5]

Country	1961	1964	1965	1966	1967
	10^9 CF				
Australia	454	2,460	4,500	4,500	7,500
Brunei-Malaysia	453	600	700	750	750
Indonesia	2,000	2,000	2,000	2,000	2,000
New Zealand	--	--	--	--	--
Philippines	1	--	--	--	--

1.2.1.9. Communist Countries

1.2.1.9.1. U.S.S.R.

By far the largest reserves of natural gas for this block are located in the Soviet Union. In 1967, 34 new gas fields and 10 fields with both oil and gas were discovered. (See Table 1. 2. 1. -9)

Large gas reserve figures, many of them conflicting, were given in various press articles during 1967 and early 1968. The most conservative, possibly giving the status at the beginning of 1967, was 140.473 trillion CF, of which 23.990 trillion CF were in the Ukraine, 24.502 in Uzbekistan, 16.707 in Turkmenistan, 2.111 in the Far East, and 73.163 in the remainder of the Soviet Union (including the North Caucasus, Volga-Ural region, and the West Siberian basin). Another report, perhaps including proved and possible reserves, gave 380 trillion CF for the Soviet Union, with 125 in the West Siberian basin. Proved, probable, and possible reserves were estimated at 2.1×10^{14} CF. Reserves of the Urengoi field of the Arctic part of the West Siberian basin were reestimated at 70-90 trillion CF, making it the largest gas field of the world. The field was reported to cover an area of 2000 sq km. Reserves of the Zapolyarnoye gas field, in the same area, were revised upward to 53 trillion CF. Additionally, large reserves of LPG are present in the gas fields of the West Siberian basin.

Table 1.2.1.-9. ESTIMATED RESERVES FOR COMMUNIST COUNTRIES

Country	Oil and Gas Journal Estimates		World Oil Estimates					IGT Estimates	
	1966	1967	1961	1964	1965	1966	1967	1966	1967
	10⁹ CF								
Albania	No estimate made	12,500	20	35	35	318	318	318	318
Bulgaria			8	1,060	1,060	1,059	883	1,060	883
Czechoslovakia			1,600	530	488	530	530	530	530
East Germany			--	500	500	500	177	500	177
Hungary			300	1,060	1,031	2,119	3,531	2,119	3,955
Poland			500	920	500	500	1,024	702	1,024
Rumania			4,800	7,770	7,316	7,500	7,062	20,000	20,000
Yugoslavia	900	3,500	125	953	923	600	847	923	3,500
U.S.S.R. (Russia)	155,000	200,000	75,000	98,374	93,812	130,647	150,000	155,000	274,000*

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*Due to increases in the reserves of Western Siberia including Tyumen, Tomsk, and Novosibirsk Provinces.

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In the Soviet Far East, a gas discovery was made at Yuzhno Sakhalinsk on the northeast coast of Sakhalin Island.

In Central Siberia, north of Irkutsk, a gas discovery was made east of Bratsk on the Ilim River. The well produced more than 17.5 million CF/day.

In the far north of Western Siberia a new gas field was found along the Messo-Yakha River, which flows into Taz Gulf in the Yenesei Lowlands. The discovery well tested at more than 17.5 million CF/day. Construction of a pipeline from this field to the Norilsk industrial center east of the Yenesei River started in 1968. A gas field, with reserves estimated at 3.5 trillion CF, was found at Nyda on the Ob Gulf.

In the province of Tomsk there were three new hydrocarbon discoveries — Olenya, Lugniezk, and Sobolina. Twelve additional seismic prospects in the province were scheduled for drilling. The Sobolina discovery well, in the populous part of the province near the city of Tomsk, produced 2100 bbl/day of good-quality oil. The first well in a new field on the Chizhapka River, a tributary of the Vasyugan, produced 2100 bbl/day of oil, and the second well tested at 17.5 million CF/day. A record gas well, producing 40 million CF/day, was completed in the Myldzinskoye field.

Four new gas fields found in Uzbekistan had reserves estimated at 3.5 trillion CF. A production center for the area is to be located at Boharain. A gas field was discovered on the Ust-Yurt Plateau at Basay, near the Aral Sea. It was being tied to the Bukhara-Ural pipeline. Another gas discovery near the Aral Sea was made at Ksiloy, from a reservoir at 460 meters.

In Turkmenistan a newly discovery gas field in the eastern Karakum Desert, Mayskoye, had estimated reserves of 950 billion CF. A gas field with a thick reservoir was found at Gugurtli in northeastern Turkmenistan. Reserves were estimated at 2.8 trillion CF. The field is to be tied to the Central Asia-Central Russia gas pipeline.

In the Don Steppe north of Rostov na Donu, the third gas field in the area was found in the Astakhov Valley near Kamensk Shakhtinskiy. The first well tested at 3.5 million CF/day.

In the Dneiper-Donetz basin the new Yefremovka and Mashevka gas fields on the left bank of the Dneiper were being tied by a line to the Dashava-

Minsk pipeline. Gas reserves of Yefremovka were estimated at 4.235 trillion CF, with 30 million bbl of condensate. Important extensions were reported to various gas fields in the Dneipropetrovsk and Poltava areas, including some significant deeper pool discoveries.

In the Volga-Ural region a multipay gas field with estimated reserves of 2.2 trillion CF in Artinskian limestones at 1200-1800 meters was discovered 20 km from Orenburg in the Oktabyr area. It was to be tied by a pipeline about 100 km long to the main gas pipeline system.

In conclusion, the Soviet Union and United States represent the two largest sources of estimated reserves of natural gas.

1.2.1.9.2. Albania

Although no reliable data exist for Albania, it appears that no new additions to reserves were made in 1967. It is of interest to note that only a small amount of gas is used. The majority of this associated gas is flared. A fertilizer plant was built to utilize the gas. (See Table 1. 2. 1.-9)

1.2.1.9.3. Bulgaria and Czechoslovakia

Neither Bulgaria or Czechoslovakia added to their reserves. These countries plan to increase imports to meet their growing demands for natural gas. (See Table 1. 2. 1.-9)

1.2.1.9.4. East Germany

1.2.1

Although exploratory drilling continues, the northeastern part of the country along the Baltic Coast, no reliable reports on results were forthcoming. (See Table 1. 2. 1.-9)

1.2.1.9.5. Hungary

In Hungary's oldest oil area, Zala County, gas was discovered at Bajesa in deeper formations. Reserves are estimated between 180 and 350 billion CF. This area may ultimately contain 3.5 trillion CF. The Tapo-Algyo oil- and gas-producing area near Szeged discovered in 1964-1965 has proved to be the largest in Hungary with gas reserves of 800 billion CF. (See Table 1. 2. 1.-9)

1.2.1.9.6. Poland

Prospecting continues in the Carpathian and Sudeta foothills, the Cracow region, Lodz, Warsaw, the Mazurian Lake district, West Pommerania, and the Lublin Plateau in Poland. A "drilling offensive" has been announced in the Carpathian area. Since development of the internal pipeline network has not been completed, the installation of gas is hampered. (See Table 1. 2. 1.-9)

1.2.1.9.7. Rumania

Exploration activity in Rumania has been at a low level with no major additions to resources.

1.2.1.9.8. Yugoslavia

In 1967, the Yugoslav-Italian boundary in the Adriatic Sea was established by treaty. Plans for offshore explorations were greatly accelerated. Yugoslavia plans to attract foreign capital for exploration and drilling. ENI and Standard Oil of Indiana have already concluded agreements. (See Table 1. 2. 1.-9)

1.2.2. Current World Gas Consumption

The data presented for consumption of natural gas during the period 1961-1967 conform, as far as possible, with the definitions given below.

Consumption is defined as production (with or without shrinkage) plus imports, minus exports, and plus or minus changes in storage. These figures, when corrected for losses (leakage, etc.), become net consumption figures. As one purpose of this study is the evaluation of the availability of gas, the difference between production and consumption figures provides a measure of the quantities of gas (produced in association with crude oil) available for liquefaction.

Tables 1.2.2.-1. through 1.2.2.-10. contain specific consumption data by area.

Table 1.2.2.-1. PRODUCTION AND CONSUMPTION FOR NORTH AMERICA (Exclusive of U.S.)

Country	Production					Consumption[c]				
	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10^9 CF									
Canada [a]	787.3	1114	1251	1125	1216	626[a]	733[a]	865[a]	716[a]	781[d]
Mexico[b]	360.5	485.0	493.1	529.2	572.8	260.0	369.4	375.8	404.5	NA
Trinidad [b]	102.3	110.7	111.5	118.9	140.3	29.3	38.5	41.3	48.7	53.5

[a] Source: Canadian Petroleum Association.

[b] Source: AAPG Bulletin.

[c] Source: UN World Energy Supplies.

[d] IGT estimate.

NA: Not available.

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Table 1.2.1.-3. PRODUCTION AND CONSUMPTION FOR SOUTH AMERICA

Country	Production[b] 1961	1964	1965	1966	1967	Consumption[e] 1961	1964	1965	1966	1967
	10⁹ CF									
Argentina	161.2	215.4	153.6	221.8	218.7	82.4	131.9	153.6	176.	Not available
Bolivia	--	8.8	8.1	12.0	30.5	--	3.1	2.8	3.2	
Brazil	18.6	18.8	24.2	27.8	30.9	1.9	2.7	2.8	2.9	
Chile	77.5	221.8	219.4	234.9	248.6	14.1	14.1	14.1	14.1	
Colombia	78.3	66.8	89.6	95.9	98.5	11.0	23.6	30.9	41.5	
Ecuador	4.7	7.1	7.1	7.1	6.1[c]	Data not available				
Peru	28.8[a]	53.8	58.1	58.9	60.1	1.7	2.4	2.5	2.6	
Venezuela	709	789	816	805	915	172.7	217.9	249.0	290	

[a] Source: UN World Energy Supplies.

[b] Source: AAPG Bulletin.

[c] Estimated.

[d] Ministerio de Minas e Hidrocarbros, Caracas.

[e] Economic Commission for Latin America.

Table 1.2.2.-3. PRODUCTION AND CONSUMPTION FOR WESTERN EUROPE

Country	Production*					Consumption ‡				
	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10^9									
Austria	55.1	62.3	60.8	66.1	63.4	55.1	62.3	60.8	66.0	Not available
France	143.8	184.1	175.9	184.3	198.0	143.8	184.1	182.6	197.7	
Germany, West	17.0	69.5	98.1†	119.7†	153.1†	17.0	69.9	99.5	121.5	
Italy	242.3‡	270.9‡	275.4	310.5	330.7	242.3	270.9	275.4	310.5	
The Netherlands	13.6	25.8	64.2	121.0	274.3	13.6	25.4	62.8	116.1	
Sicily	Included under Italy									
Spain	--	--	0.13	0.13	0.13	--	0.16	0.16	0.13	
United Kingdom	2.8‡	6.3†	6.7†	6.3†	17.00	2.8	9.5	29.3	27.7	

*Source: AAPG Bulletin.

†Includes German share of gas from Groningen field produced from wells in The Netherlands.

‡Source: UN World Energy Supplies.

Table 1.2.2.-4. PRODUCTION AND CONSUMPTION FOR NORTH AFRICA

	Production[*]					Consumption				
Country	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10^9 CF									
Algeria	27.8	28.6	61.8	69.1	113.0	10.2[†]	25.4[†]	35.3[†]	37.1[†]	NA
Egypt	--	10.3	10.6	9.7	15.0	Negligible gas consumption				
Libya	4.3	219.6	306.6	356.0	637.0	All flared				
Morocco	0.32	0.37	0.40	0.40	0.33	0.32	0.37	0.40	0.40	0.33
Tunisia	0.26	0.28	0.30	0.30	0.33	0.26	0.28	0.30	0.30	0.33

[*] Source: AAPG Bulletin.

[†] Source: UN World Energy Supplies.

NA: Not available.

Table 1.2.2.-5. PRODUCTION AND CONSUMPTION FOR WEST AFRICA

Country	Production*					Consumption†				
	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10⁹ CF									
Angola	0.165	0.300	0.414	0.701	0.226	Not available				
Gabon	0.237	0.334	0.376	0.406	0.615	0.35	0.35	0.35	0.35	NA
Nigeria	--	36.3	79.1	103.7	94.2	--	1.8	3.5	6.4	NA

* Source: AAPG Bulletin.

† Source: UN World Energy Supplies.

NA: Not available.

Table 1.2.2.-6. PRODUCTION AND CONSUMPTION FOR MIDDLE EAST

Country	Production* 1961	1964	1965	1966	1967	Consumption† 1961	1964	1965	1966	1967
	10⁹									
Abu Dhabi	--	27	40	52	55	Gas flared				
Bahrain	6	7	8	9	10	Small quantities used in refinery supply and power generation				
Iran	208	293	326	370	450	34.7	42.0	43.4	49.0	NA
Iraq	144	180	187	198	183	22.4	26.9	27.2	31.8	NA
Israel	--	1.1†	2.5†	3.2†	4.0	0.16	0.95	2.5	3.2	NA
Kuwait	343	443	453	475	478	38.3	59.2	63.3	89.7	135.8
Neutral Zone	12	24	24	28	28	Data not available				
Omar	No production				9					
Qatar	25	31	33	42	46	Data not available				
Saudi Arabia	175	216	254	300	326	--	0.07	0.07	1.06	NA
Syria	No reliable data available									
Turkey	1	2.5	4	6	8	1	2.5	4	6	NA

* IGT estimate, including flared and reinjected gas.

† Source: UN World Energy Supplies.

NA: Not available.

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Table 1.2.2.-7. PRODUCTION AND CONSUMPTION FOR FAR EAST

	Production*					Consumption*				
	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
Country	10⁹ CF									
Afghanistan	Gas production negligible									
Burma	--	3.1	3.1	2.9	2.9	0.35†	0.35†	0.35†	0.35†	NA
China	No reliable data available									
India	--	8.3	10.95	10.95	18.3	--	8.3	10.95	10.95	18.3
Japan	33.6	58.3	62.9	64.5	50.9	33.6	65.0	62.9	64.5	50.9
Pakistan	27.1	50.2	60.9	70.5	77.6	27.1	50.2	60.9	70.5	77.6
Taiwan	1.2	6.0	10.9	15.7	18.7	1.2	6.0	10.8	15.7	18.7

* Source: AAPG Bulletin.

† Source: UN World Energy Supplies.

NA: Not available.

Table 1.2.2.-8. PRODUCTION AND CONSUMPTION FOR OCEANIA

	Production					Consumption				
Country	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10^9 CF									
Australia*	--	0.468	0.951	1.174	--	Gas consumption neglible				
Brunei-Malaysia†	2.83	6.0	6.0	7.1	--	No gas consumption				
Indonesia	18.2	96.4	86.5	56.5		Gas consumption negligible				
New Zealand	No production recorded					No gas consumption				

* Source: AAPG Bulletin.

† Source: UN World Energy Supplies.

Table 1.2.2.-9. PRODUCTION AND CONSUMPTION FOR COMMUNIST COUNTRIES

Country	Production† 1961	Production† 1964	Production† 1965	Production† 1966	Production† 1967	Consumption* 1961	Consumption* 1964	Consumption* 1965	Consumption* 1966	Consumption* 1967
	10⁹ CF									
Albania	Not available									
Bulgaria	--	--	1.8	2.54	10.2	--	--	1.8	2.5	Not available
Czechoslovakia *	44.5	32.8	30.0	45.9	--	44.5	32.8	30.0	45.9	
Germany, East *	0.71	3.9	0.46	4.2	--	0.71	3.9	0.46	4.2	
Hungary	11.4	27.7	39.1	54.8	72.2	18.7	35.7	46.3	61.8	
Poland	12.2	43.5	48.1	48.6	55.5	35.7	54.0	62.2	73.5	
Rumania	255.0	403.0	454.2	497.0	566.0	247.9	395.6	447.1	489.9	
Yugoslavia	2.4	9.5	12.0	14.2	21.4	2.4	9.5	12.0	14.2	
U.S.S.R.	2150.0	3753.0	4553.0	5118.5	5612.7	2140.5	3742.8	4539.2	5089.2	

* Source: UN World Energy Supplies.

† Source: AAPG Bulletin.

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Table 1.2.2.-10. PRODUCTION AND CONSUMPTION FOR UNITED STATES

	Production*					Consumption†				
	1961	1964	1965	1966	1967	1961	1964	1965	1966	1967
	10^9 CF									
United States	13,379	15,347	16,282	17,491	18,381	3587	15,761	16,682	17,945	18,841

* Net marketed production, source: American Gas Association.

† Includes imported natural gas, source: American Gas Association.

1.2.3. Current World Gas Production

The data presented for production of natural gas during the period 1961-1967 conform, as far as possible, with the definitions given below.

Production is defined as the total volume of natural gas withdrawn from producing reservoirs less the volume returned to such reservoirs in cycling, repressuring of oil reservoirs, and conservation operations. As used, net production is production corrected for shrinkage resulting from the removal of liquefiable portions of the gas. Data for the United States and Canada are on a net production basis. For other countries the basis is generally production.

Tables 1.2.2.-1. through 1.2.2.-10. contain specific production data by area.

1.2.3.1. Canada

Table 1.2.3.-1 gives the world LNG production facilities in operation or planned.

Table 1.2.3.-1. CANADIAN LNG PRODUCTION FACILITIES

Location	Liquefaction Capacity, 10^6 CF/day	Use
Northern and Central Gas Co., Ltd., Sudbury, Ontario	2.5	Peakshaving
Cryogenic Enterprises, Richmond, B. C.	0.4	Remote supply
Vancouver, B. C.	4.3	Peakshaving
Quebec Natural Gas Corp., Montreal, Quebec	10.0	Peakshaving

1.2.3.2. Other Countries

The facilities for liquefying natural gas for export presently known about are given in Table 1.2.3.-2.

Table 1.2.3.-2. WORLD LNG PRODUCTION FACILITIES OUTSIDE OF CANADA AND THE U.S.

Location	Liquefaction Capacity, 10^6 CF/day	Status
Port Arzew, Algeria	145	Operating
Marsa el Brega, Libya	345	Final testing
Cook Inlet, Alaska	140	Construction
Skikda, Algeria	435	Final planning
Pozo Rica, Mexico	275	Planning
Maricaibo, Venezuela	--	Planning
Brunei, Malaysia	200	Final planning

1.2.4. Probable World Consumption/Production

The probable world consumption and production of natural gas through 1985 is detailed for each region in Tables 1.2.4.-1 through 1.2.4.-9. (Section 1.2.4.1. to Section 1.2.4.9).

1.2.4.1. North and Central America (Excluding the U.S.)

Table 1.2.4.-1. PROJECTED PRODUCTION AND CONSUMPTION FOR NORTH AMERICA
(Exclusive of U.S.)

	Production				Consumption			
	1970	1975	1980	1985	1970	1975	1980	1985
Country	10^9 CF							
Canada	1860	2577	3496	4875	1248	1664	2146	2865
Mexico	581	687	813	920	442	582	755	880
Trinidad*	182	246	332	448	71	96	129	175

* Based on recent unannounced discoveries.

1.2.4.2. South America

Table 1.2.4.-2. PROJECTED PRODUCTION AND CONSUMPTION FOR SOUTH AMERICA

	Production				Consumption			
	1970	1975	1980	1985	1970	1975	1980	1985
Country	10^9 CF							
Argentina	425	510	590	670	425	550	700	900
Bolivia	35.5	393.1	317.3	328.8	11.8	17.6	23.0	29.3
Brazil	40.1	44.3	48.9	54.0	3.7	5.2	6.7	8.8
Chile	271.7	314.9	365.0	423.0	28.2	35.3	42	49
Colombia	107.7	118.9	144.7	176.1			81.2	103
Ecuador	7.1	7.5	7.9	9.2				
Peru	63.8	70.4	77.7	85.8	3.0	4.0	5.2	6.5
Venezuela	862	819	1036	1260	364	523	742	960

1.2.4.3. Western Europe

Table 1.2.4.-3. PROJECTED PRODUCTION AND CONSUMPTION FOR WESTERN EUROPE

	Production				Consumption			
Country	1970	1975	1980	1985	1970	1975	1980	1985
	10⁹ CF							
Austria	55	42	34	27	78	96	120	150
Belgium	No production, imports from The Netherlands				170	197	230	266
France	229	353	453	525	505	694	945	1266
Germany, West	188	505	645	747	495	830	1132	1515
Italy	383	591	820	1046	488	606	1091	1460
The Netherlands	985	1421	1883	2225	250	722	1011	1353
Sicily	Included under Italy							
Spain	No production, imports LNG				44	135	195	245
United Kingdom	531.5	893.5	1365	1848	568	930.5	1425	1908

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1.2.4.4. North Africa

Table 1.2.4.-4 PROJECTED PRODUCTION AND CONSUMPTION FOR NORTH AFRICA

Country	Production 1970	Production 1975	Production 1980	Production 1985	Consumption 1970	Consumption 1975	Consumption 1980	Consumption 1985
	10⁹ CF							
Algeria	240	367	537	761	20	30.5	44	63.5
Egypt	60	92	134	190	60	92	134	190
Libya	643	886	1045	1218	14.6	34.8	47.6	80.0
Morocco	0.33	0.50	0.73	1.04	0.33	0.50	0.73	1.04
Tunisia	0.44	0.67	0.98	1.4	0.44	0.67	0.98	1.4

1.2.4.5 West Africa

Table 1.2.4.-5. PROJECTED PRODUCTION AND CONSUMPTION FOR WEST AFRICA

Country	Production 1970	Production 1975	Production 1980	Production 1985	Consumption 1970	Consumption 1975	Consumption 1980	Consumption 1985
	10⁹ CF							
Angola	0.90	1.38	2.02	2.86	Used as fuel gas in field and refineries, remainder flared			
Gabon	1.2	1.8	2.6	3.7				
Nigeria	405	620	906	1283	20	32	45	64

8926

1.2.4.6. Middle East

Table 1.2.4.-6. PROJECTED PRODUCTION AND CONSUMPTION FOR MIDDLE EAST

	Production				Consumption			
Country	1970	1975	1980	1985	1970	1975	1980	1985
	10⁹ CF							
Abu Dhabi	61.9	75.3	91.7	111.6	Negligible consumption			
Bahrain	11.3	13.8	16.7	20.4	Negligible consumption			
Iran	447	730	1050	1237	230	380	550	635
Iraq	236	286	336	386	67	89.2	115	146
Israel	4.5	5.5	6.7	8.1	4.5	10.4	16.7	28.1
Kuwait	520	710	830	950	168	222	285	358
Neutral Zone	31.5	38.3	46.7	56.8	No present or projected plans for consumption outside field use; reinjection attempt at Dukhan reservoir in Qatar unsuccessful.			
Oman	10.1	12.3	15.6	18.2				
Qatar	51.8	63.0	76.7	93.4				
Saudi Arabia	340	426	512	598	97	133	175	225
Turkey	9.0	11.0	13.4	16.2	9.0	11.0	13.4	16.2

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1.2.4.7. Far East

Table 1.2.4.-7. PROJECTED PRODUCTION AND CONSUMPTION FOR FAR EAST

	Production				Consumption			
	1970	1975	1980	1985	1970	1975	1980	1985
Country	10^9 CF							
Afghanistan	110	115	118	120	--	5	8	10
India	20.6	25.1	30.5	37.2	20.6	25.1	30.5	37.2
Japan	74	85	96	109	123	206	350	600
Pakistan	100	165	280	465	100	165	280	465
Taiwan	40.8	41.3	41.3	41.3	40.8	41.3	41.3	41.3

1.2.4.8. Oceania

Table 1.2.4.-8. PROJECTED PRODUCTION AND CONSUMPTION FOR OCEANIA

	Production				Consumption			
	1970	1975	1980	1985	1970	1975	1980	1985
Country	10^9 CF							
Australia	90	236	331	464	90	236	331	464
Brunei, Malaysia	53.5	65.1	79.2	96.4	2.7*	3.3	4.0	4.8
Indonesia	178	216	263	320	9†	11	13	16
New Zealand	14.6	18.5	23.4	29.6	14.6	18.5	23.4	29.6

* Oil field consumption; 7.5 X 10^9 CF exported as LNG to Japan for period 1971-1991.

† Oil field consumption.

1.2.4.9. Communist Countries

Table 1.2.4.-9. PROJECTED PRODUCTION AND CONSUMPTION FOR COMMUNIST COUNTRIES

	Production				Consumption			
	1970	1975	1980	1985	1970	1975	1980	1985
Country	10⁹ CF							
Albania	Insufficient data for projection							
Bulgaria	24.7	46.5	70.6	112	24.7	46.5	70.6	112
Czechoslovakia	38.7	33.7	25.2	18.4	74.1	98	119	143
Germany, East	No significant foreseeable production				Imports probable but quantities not yet predictable			
Hungary	117	145	160	172	117	145	160	172
Poland	96	155	195	215	150	260	335	395
Rumania	718	940	1160	1390	718	940	1160	1390
Yugoslavia	35	70	110	175	35	100	150	250
U.S.S.R. (Russia)	7600	12,500	24,500	49,000	7700	12,500	24,000	48,600

1.2.5. Availability of Gas for LNG Production/Export

Tables 1.2.5.-1 through 1.2.5.-9 details the availability of natural gas for LNG production and export.

Table 1.2.5.-1. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN NORTH AND CENTRAL AMERICA
(Excluding U.S.)

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Canada	612	913	1350	2010
Mexico	139	105	58	40
Trinidad	111	150	203	273
Total	862	1168	1611	2323

Table 1.2.5.-2. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN SOUTH AMERICA

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Argentina	0	0	0	0
Bolivia	23.7	375.5	294.3	299.5
Brazil	36.4	39.1	42.2	45.2
Chile	243.5	279.6	323.0	374.0
Colombia	66.4	58.9	63.5	73.1
Ecuador	0	0	0	0
Peru	60.8	66.4	72.5	79.3
Venezuela	498	296	294	300
Total	922.8	1115.5	1089.5	1171.1

Table 1.2.5.-3. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN WESTERN EUROPE

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Austria	No gas available for export, primary market for LNG.			
Belgium				
France				
Germany, West				
Italy				
The Netherlands				
Sicily				
Spain				
United Kingdom				

Table 1.2.5.-4. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN NORTH AFRICA

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Algeria	220	336	493	697
Egypt	0	0	0	0
Libya	628	851	997	1138
Morocco	0	0	0	0
Tunisia	0	0	0	0
Total	848	1187	1490	1835

Table 1.2.5.-5. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN WEST AFRICA

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Angola	0.9	1.38	2.0	2.86
Gabon	1.2	1.8	2.6	3.7
Nigeria	385	588	861	1219
Total	387.1	591.18	865.6	1225.56

Table 1.2.5.-6. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN MIDDLE EAST

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Abu Dhabi	61.9	75.3	91.7	111.6
Bahrain	11.3	13.8	16.7	20.4
Iran	117	350	500	602
Iraq	169	197	221	240
Israel	0	0	0	0
Kuwait	422	488	555	597
Neutral Zone	31.5	38.3	46.7	56.8
Oman	10.11	12.3	15.0	18.2
Qatar	51.8	63.0	76.7	93.4
Saudi Arabia	243	293	337	373
Turkey	0	0	0	0
Total	1117.61	1530.7	1859.8	2112.4

Table 1.2.5.-7. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN FAR EAST

Country	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
Afghanistan	110	110	110	110
India	0	0	0	0
Japan	0	0	0	0
Pakistan	0	0	0	0
Taiwan	0	0	0	0
Total	110	110	110	110

Table 1.2.5.-8.. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN OCEANIA

	1970	1975	1980	1985
Country	Available Gas, 10^9 CF/yr			
Australia	0	0	0	0
Brunei, Malaysia	50.8	61.8	75.2	91.6
Indonesia	169	205	250	304
New Zealand	0	0	0	0
Total	219.8	266.8	325.2	395.6

Table 1.2.5.-9. AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT IN COMMUNIST COUNTRIES

	1970	1975	1980	1985
Country	Available Gas, 10^9 CF/yr			
Albania	0	0	0	0
Bulgaria	0	0	0	0
Czechoslovakia	0	0	0	0
Germany, East	0	0	0	0
Hungary	0	0	0	0
Poland	0	0	0	0
Rumania	0	0	0	0
Yugoslavia	0	0	0	0
U. S. S. R. (Russia)	0	0	500	400
Total	0	0	500	400

Table 1.2.5.-10. TOTAL WORLD AVAILABILITY OF GAS FOR LNG PRODUCTION/EXPORT

Area	1970	1975	1980	1985
	Available Gas, 10^9 CF/yr			
North and Central America (excluding U.S.)	862	1168	1611	2323
South America	922.8	1115.5	1089.5	1171.1
Western Europe	0	0	0	0
North Africa	848	1187	1490	1835
West Africa	387.1	591.18	865.6	1225.56
Middle East	1117.61	1530.7	1859.8	2112.4
Far East	110	110	110	110
Oceania	219.8	266.8	325.2	395.6
Communist Countries	0	0	500	400
Total	4467.31	5969.18	7851.1	9572.66

1. 3. United States/World Availability Ratios

1. 3. 1. Natural Gas Availability Ratios

The ratios of natural gas production to consumption were computed for each region of the world. These ratios allow a rapid screening of those regions and countries that have an exportable surplus of natural gas. Coupled with the consumption and production estimates, these ratios allow direct evaluation of potential sources of supply and the amount of that supply. Conversely, the ratios provide an indication of those regions and countries that are deficient in natural gas. The data are summarized in Tables 1. 3. 1. -1 through 1. 3. 1. -9.

Table 1.3.1.-1. NATURAL GAS AVAILABILITY RATIO FOR NORTH AND CENTRAL AMERICA (excluding U. S.)

	1970	1975	1980	1985
Country	Availability Ratio			
Canada	1. 49	1. 55	1. 63	1. 70
Mexico	1. 31	1. 18	1. 08	1. 05
Trinidad	2. 56	2. 56	2. 57	2. 56

Table 1.3.1.-2. NATURAL GAS AVAILABILITY RATIOS FOR SOUTH AMERICA

	1970	1975	1980	1985
Country	Availability Ratio			
Argentina	1. 0	0. 93	0. 84	0. 74
Bolivia	30. 1	22. 34	13. 80	11. 22
Brazil	10. 84	8. 52	7. 30	6. 14
Chile	9. 63	8. 92	8. 69	8. 63
Colombia	2. 61	1. 98	1. 78	1. 71
Ecuador	1. 0	1. 0	1. 0	1. 0
Peru	21. 3	17. 6	14. 9	13. 2
Venezuela	2. 37	1. 57	1. 40	1. 31

Table 1.3.1.-3. NATURAL GAS AVAILABILITY RATIOS FOR WESTERN EUROPE

Country	1970	1975	1980	1985
	Availability Ratio			
Austria	0.71	0.44	0.28	0.18
Belgium	No production, fully dependent on imports			
France	0.45	0.51	0.48	0.41
Germany, West	0.38	0.61	0.57	0.49
Italy (including Sicily)	0.78	0.85	0.75	0.72
The Netherlands	3.94	1.97	1.86	1.64
Spain	No production, fully dependent on imports			
United Kingdom	0.94	0.96	0.96	0.97

Table 1.3.1.-4. NATURAL GAS AVAILABILITY RATIOS FOR NORTH AFRICA

Country	1970	1975	1980	1985
	Availability Ratio			
Algeria	12.0	12.0	12.2	12.0
Egypt	1.0	1.0	1.0	1.0
Libya	44.0	25.5	22.0	15.2
Morocco	1.0	1.0	1.0	1.0
Tunisia	1.0	1.0	1.0	1.0

Country	1970	1975	1980	1985
	Availability Ratio			
Angola Gabon	Minimal internal consumption; therefore, ratios would be meaningless.			
Nigeria	20.3	19.4	20.1	20.0

Table 1.3.1.-6. NATURAL GAS AVAILABILITY RATIOS FOR MIDDLE EAST

	1970	1975	1980	1985
Country	Availability Ratio			
Abu Dhabi	Minimal internal consumption; therefore, ratios would be meaningless.			
Bahrain				
Iran	1.94	1.92	1.91	1.95
Iraq	3.52	3.21	2.92	2.64
Israel	1.0	0.53	0.40	0.29
Kuwait	3.52	3.20	2.91	2.65
Neutral Zone	No present reliable information on expected consumption.			
Oman				
Qatar				
Saudi Arabia	3.50	3.20	2.93	2.66
Turkey	1.0	1.0	1.0	1.0

Table 1.3.1.-7. NATURAL GAS AVAILABILITY RATIOS FOR FAR EAST

	1970	1975	1980	1985
Country	Availability Ratio			
Afghanistan	11.0	23.0	14.75	12.0
India	1.0	1.0	1.0	1.0
Japan	0.60	0.41	0.27	0.18
Pakistan	1.0	1.0	1.0	1.0
Taiwan	1.0	1.0	1.0	1.0

Table 1.3.1.-8. NATURAL GAS AVAILABILITY RATIOS FOR OCEANIA

Country	1970	1975	1980	1985
	Availability Ratios			
Australia	1.0	1.0	1.0	1.0
Brunei, Malaysia	19.8	19.7	19.8	20.0
Indonesia	19.7	24.2	20.2	20.0
New Zealand	1.0	1.0	1.0	1.0

Table 1.3.1.-9. NATURAL GAS AVAILABILITY RATIOS FOR COMMUNIST COUNTRIES

Country	1970	1975	1980	1985
	Availability Ratios			
Albania	Insufficient data			
Bulgaria	1.0	1.0	1.0	1.0
Czechoslovakia	0.52	0.34	0.21	0.13
Germany, East	No significant production			
Hungary	1.0	1.0	1.0	1.0
Poland	0.64	0.59	0.58	0.54
Rumania	1.0	1.0	1.0	1.0
Yugoslavia	1.0	0.70	0.73	0.7
U.S.S.R. (Russia)	0.99	1.0	1.02	1.01

1.3.2. LNG Availability Ratios

It is unrealistic to try and define LNG availability ratios at the present time. In the United States, the major use of LNG is for peakshaving. Very little is available for other uses. As for world availability, since only two plants are presently operating with the rest in the early stages of planning, it is assumed that the availability of LNG worldwide will keep pace with demand.

1.4. Projected Energy Requirements

1.4.1. Coal Equivalents

The estimates of world natural gas consumption were converted to coal equivalents on the basis of 13,000 Btu/lb of coal and 1000 Btu/SCF of natural gas. These coal equivalents are presented for each region in Tables 1.4.1.-1 through 1.4.1.-9.

Table 1.4.1.-1. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR NORTH AND CENTRAL AMERICA (exclusive of U.S.)

Country	1970	1975	1980	1985
	— Energy Requirements, 10^6 tons coal equivalent —			
Canada	48.0	64.0	82.5	110
Mexico	17.0	22.4	29.0	33.1
Trinidad	2.73	3.69	4.96	6.73

Table 1.4.1.-2. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR SOUTH AMERICA

	1970	1975	1980	1985
Country	Energy Requirements, 10^6 tons coal equivalent			
Argentina	16.3	21.2	26.9	34.6
Bolivia	0.45	0.68	0.88	1.13
Brazil	0.14	0.20	0.26	0.34
Chile	1.08	1.36	1.62	1.88
Colombia	1.59	2.31	3.12	3.96
Ecuador	Minimal usage in field			
Peru	0.12	0.15	0.20	0.25
Venezuela	14.0	20.1	28.5	36.9

Table 1.4.1.-3. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR WESTERN EUROPE

	1970	1975	1980	1985
Country	Energy Requirements, 10^6 tons coal equivalent			
Austria	3.0	3.70	3.85	5.77
Belgium	6.54	7.58	8.85	10.2
France	19.4	26.7	36.3	48.7
Germany, West	19.0	31.9	43.5	58.3
Italy, including Sicily	18.8	26.8	42.0	56.2
The Netherlands	9.62	27.8	38.9	52.0
Spain	1.69	5.19	7.50	9.42
United Kingdom	21.8	35.0	54.8	73.4

Table 1.4.1.-4. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR NORTH AFRICA

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 tons coal equivalent —			
Algeria	0.77	1.17	2.69	3.44
Egypt	2.31	3.54	5.15	7.31
Libya	0.56	1.34	1.83	3.08
Morocco	0.01	0.02	0.03	0.04
Tunisia	0.02	0.03	0.04	0.05

Table 1.4.1.-5. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR WEST AFRICA

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 tons coal equivalent —			
Angola	0	0	0	0
Gabon	— Minimal use in oil fields —			
Nigeria	0.77	1.23	1.73	2.46

Table 1.4.1.-6. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR MIDDLE EAST

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 tons coal equivalent —			
Abu Dhabi Bahrain	— Negligible consumption —			
Iran	8.85	14.6	21.2	24.4
Iraq	2.57	3.43	4.42	5.62
Israel	0.17	0.40	0.64	1.08
Kuwait	6.46	7.77	11.0	13.8
Neutral Zone Oman Qatar	No consumption other than minimal field usage			
Saudi Arabia	3.73	5.12	6.73	8.65
Turkey	0.35	0.42	0.52	0.62

Table 1.4.1.-7. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR FAR EAST

	1970	1975	1980	1985
Country	Energy Requirements, 10^6 tons coal equivalent			
Afghanistan	0	0.20	0.31	0.38
India	0.79	0.97	1.18	1.43
Japan	4.73	7.92	13.5	23.1
Pakistan	3.85	6.34	10.8	17.9
Taiwan	1.57	1.59	1.59	1.59

Table 1.4.1.-8. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR OCEANIA

	1970	1975	1980	1985
Country	Energy Requirements, 10^6 tons coal equivalent			
Australia	2.69	9.08	12.8	17.8
Brunei, Malaysia	0.10	0.13	0.15	0.18
Indonesia	0.34	0.42	0.50	0.62
New Zealand	0.56	0.71	0.90	1.14

Table 1.4.1.-9. PROJECTED ENERGY REQUIREMENTS IN COAL EQUIVALENTS FOR COMMUNIST COUNTRIES

	1970	1975	1980	1985
Country	Energy Requirements, 10^6 tons coal equivalent			
Albania	Insufficient data			
Bulgaria	0.95	1.79	2.72	4.31
Czechoslovakia	2.85	3.77	4.58	5.50
Germany, East	Insufficient data			
Hungary	4.50	5.58	6.15	6.62
Poland	5.77	10.0	12.9	15.2
Rumania	27.6	36.2	44.6	53.5
Yugoslavia	1.35	3.85	5.77	9.62
U.S.S.R. (Russia)	296	481	923	1870

1.4.2. Oil Equivalents

The estimates of world natural gas consumption were converted to oil equivalents on the basis of 6.1 million Btu/bbl of oil and 1000 Btu/SCF of natural gas. The oil equivalents are presented for each region in Tables 1.4.2.-1 through 1.4.2.-9.

Table 1.4.2.-1. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR NORTH AND CENTRAL AMERICA
(Exclusive of U.S.)

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 bbl oil equivalent —			
Canada	204	273	351	469
Mexico	72.4	95.4	124	141
Trinidad	11.6	15.7	21.1	28.7

Table 1.4.2.-2. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR SOUTH AMERICA

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 bbl oil equivalent —			
Argentina	69.4	90.3	115	147
Bolivia	1.92	2.90	3.75	4.81
Brazil	0.60	0.85	1.10	1.45
Chile	4.60	5.79	6.90	8.01
Colombia	6.77	9.84	13.3	16.9
Ecuador	— Minimal usage in oil fields —			
Peru	0.51	0.64	0.85	1.06
Venezuela.	59.6	85.6	121	157

Table 1.4.2.-3. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR WESTERN EUROPE

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 bbl oil equivalent —			
Austria	12.8	15.8	16.4	24.6
Belgium	27.9	32.3	37.7	43.5
France	82.6	114	155	207
Germany, West	80.9	136	185	248
Italy, including Sicily	80.1	114	179	239
The Netherlands	41.0	118	166	222
Spain	7.20	22.1	32.0	40.1
United Kingdom	93.0	149	233	313

Table 1.4.2.-4. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR NORTH AFRICA

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 bbl oil equivalent —			
Algeria	3.28	4.98	11.5	14.7
Egypt	9.84	15.1	21.9	31.1
Libya	2.39	5.71	7.80	13.1
Morocco	0.05	0.08	0.13	0.17
Tunisia	0.08	0.13	0.17	0.21

Table 1.4.2.-5. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR WEST AFRICA

	1970	1975	1980	1985
Country	— Energy Requirements, 10^6 bbl oil equivalent —			
Angola Gabon	— Minimal usage in oil fields —			
Nigeria	3.28	5.24	7.37	10.5

Table 1.4.2.-6. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR MIDDLE EAST

Country	1970	1975	1980	1985
	Energy Requirements, 10^6 bbl oil equivalent			
Abu Dhabi Bahrain	Negligible consumption			
Iran	37.7	62.2	90.3	104
Iraq	10.9	14.6	18.8	23.9
Israel	0.72	1.70	2.72	4.6
Kuwait	27.5	33.1	46.9	58.8
Neutral Zone Oman Qatar	No consumption other than minimal oil field usage			
Saudi Arabia	15.9	21.8	28.7	36.8
Turkey	1.49	1.79	2.22	2.64

Table 1.4.2.-7. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR FAR EAST

Country	1970	1975	1980	1985
	Energy Requirements, 10^6 bbl oil equivalent			
Afghanistan	0	0.85	1.32	1.62
India	3.37	4.13	5.03	6.09
Japan	20.1	33.7	57.5	98.4
Pakistan	16.4	27.0	46.0	76.3
Taiwan	6.69	6.77	6.77	6.77

Table 1.4.2.-8. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR OCEANIA

Country	1970	1975	1980	1985
	Energy Requirements, 10^6 bbl oil equivalent			
Australia	11.5	38.7	54.5	75.8
Brunei, Malaysia	0.43	0.55	0.64	0.77
Indonesia	1.45	1.79	2.13	2.64
New Zealand	2.39	3.02	3.83	4.86

Table 1.4.2.-9. PROJECTED ENERGY REQUIREMENTS IN OIL EQUIVALENTS FOR COMMUNIST COUNTRIES

Country	1970	1975	1980	1985
	Energy Requirements, 10^6 bbl oil equivalent			
Albania	Insufficient data			
Bulgaria	4.05	7.63	11.6	18.4
Czechoslovakia	12.1	16.1	19.5	23.4
Germany, East	Insufficient data			
Hungary	19.2	23.8	26.2	28.2
Poland	24.6	42.6	55.0	64.8
Rumania	118	154	190	228
Yugoslavia	5.75	16.4	24.6	41.0
U.S.S.R.(Russia)	1260	2050	3930	1970

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1. American Gas Association, 1968 Gas Facts: A Statistical Record of the Gas Utility Industry in 1966. New York, 1967.
2. American Gas Association, American Petroleum Institute and Canadian Petroleum Association, Reserves of Crude Oil. Natural Gas Liquids, and Natural Gas in the United States and Canada as of December 31, 1967. 22, New York, 1968.
3. Broadhurst, F., "Off-Cape Oil Survey Team Awaits Permit for Probe," Cape Cod Standard-Times 32 (1968) June 17.
4. Linden, H. R., "Revised Forecast of U. S. Natural Gas Supply." Paper presented to the Natural Gas Men of Houston, Houston, December 10, 1968.
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Section 2.0. Engineering Review and General Appraisal of LNG Liquefaction, Transportation, Storage, and Vaporization

The technique of liquefying and handling gases dates back more than 100 years to the time when advances were made in cryogenic technology. However, the first reported interest in liquefying natural gas occurred during World War I when the British Government proposed to build a facility that was primarily for the recovery of helium, but that concurrently liquefied natural gas. During the late 1920's and the early 1930's, development continued, with the United States Bureau of Mines building a helium recovery plant in Texas. However, the first facility primarily devoted to liquefying and storing natural gas was a pilot peakshaving plant built by the Hope Natural Gas Co. in 1940. The liquefaction plant, erected in West Virginia, had a capacity of 300,000 CF/day and a storage tank with a capacity of 1.2 million CF. Subsequently, in 1941, the East Ohio Gas Co., a sister company of the Hope Natural Gas Co., built a peakshaving plant in Cleveland, Ohio. The liquefaction plant capacity was 4 million CF/day. The storage system consisted of three spherical tanks, each with a capacity of 56 million CF of gas. These tanks had an inner shell made of 3.5 % nickel steel which was 57 ft in diameter. The inner steel shell was surrounded with cork insulation. The outer shell — 63 ft in diameter — was made of carbon steel.

The plant had successfully operated for 2-1/2 years when a new storage tank with a capacity of 100 million CF was added. This latter tank was constructed of similar materials, except that the insulation was rock wool and the top and bottom heads of the tank were somewhat dished. In 1944 this tank failed, resulting in a disastrous fire. The catastrophe occurred because the lack of suitable diking allowed the liquid to flood a large area, where it ignited. A report of the investigation into causes of the accident, conducted by the Bureau of Mines, was published and cited several possible causes for the disaster, the chief of these being the use of 3.5% nickel

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steel for the inner shell. This alloy was proved to be unsuitable for storage of LNG at −260°F.

In 1947 the United States, under a lease arrangement with the U. S. S. R., built a natural gas liquefaction plant near Moscow. This plant had a liquefaction capacity of 4.25 million CF/day and a storage capacity of 162 million CF. Subsequently, the Russians increased the plant capacity to approximately 20 million CF/day. Five years later the Union Stockyard and Transit Co. of Chicago started investigations into the feasibility of transporting large quantities of LNG from the U. S. Gulf Coast to Chicago up the Mississippi River on specially constructed barges. In 1955 Constock, Inc. (a joint venture of Union Stockyards and Continental Oil Co.), began a large research program to investigate liquefaction and barge transport of LNG. A year later a 7 million CF/day, barge-mounted expander liquefaction plant was constructed at Lake Charles, La., for liquefying gas for overseas shipment. In 1959 the Methane Pioneer, a specially reconstructed tanker, began successful shipments of LNG from Lake Charles, La., to Canvey Island, England. Storage facilities at Lake Charles consisted of double-walled tanks with an aluminum inner shell and a steel outer shell. Tank capacity was 120 million CF. A cryogenic in-ground storage container was also built.

In 1960 the American Gas Association undertook research on the belowground storage of LNG in insulated prestressed concrete tanks at the Institute of Gas Technology. Simultaneously, other groups began investigations of novel low-cost methods of LNG storage. As a result of these efforts, four types of storage system are now available to the gas industry for liquefied natural gas.

In 1961, the British Government approved a plan for liquefying natural gas at Arzew, Algeria, and shipping it in two tankers to Canvey, England, where it would be regasified and fed into the gas grid. In 1964, the first commercial shipments of LNG were made from Algeria to the United Kingdom and France as part of the overall LNG project. This project has successfully operated since then, delivering approximately 100 billion CF of gas to the United Kingdom.

The present state of the art and expected future trends in gas purification, liquefaction, storage, transportation, and vaporization are outlined and discussed. Detailed investment and operating costs are provided for most processes and systems covered. In instances where cost data are not available because the process or system is still undergoing development or because it was never tested under full-scale conditions, a qualitative evaluation has been included. It should be pointed out that much of the cost data presented in Section 2 are for specific plants or installations and are not directly comparable with the optimized cost estimates used in preparing the cost of imported LNG in Section 3.

2. 1. Current Technology

2. 1. 1. Gas Purification Processes

Before natural gas can be liquefied it must be purified; that is, the small traces of contaminating gases that might solidify on cooling and plug the piping or foul heat exchangers must be removed. Regardless of the type of liquefaction cycle used, as a general rule, the natural gas entering the liquefaction plant has to be treated for the removal of carbon dioxide, water, sulfur compounds, and other natural gas constituents that could form solids when the gas is liquefied. Removal of water is accomplished by dehydration with a dry dessicant or glycol. Dessicants commonly used are activated alumina, activated oxide, silica gel, and molecular sieves. Removal of carbon dioxide and hydrogen sulfide is accomplished by commercial gas-sweetening operatings, which are described in detail in the following sections.

2. 1. 1. 1. Dehumidification

Water vapor is probably the commonest undesirable impurity in natural gas streams. Usually, it is not the water vapor itself that is objectionable but rather the liquid or solid phase that may precipitate from the gas when it is compressed or cooled. Liquid water almost always accelerates corrosion, and ice or solid hydrates can plug valves, fittings, and even gas lines. The quantity of water in saturated natural gas at various pressures and temperatures can be estimated from Figure 2.1.1. -1. This chart also shows the hydrate formation line for 0. 6 specific gravity gas.[16] To the left of this line, solid hydrates will form when saturated gas is cooled. For example, if 1000-psia gas containing 19 lb water/million CF of gas (0.6 specific gravity) is cooled to 60°F, the solid hydrate will form. As obtained from pipelines, a gas hydrate looks like packed snow.

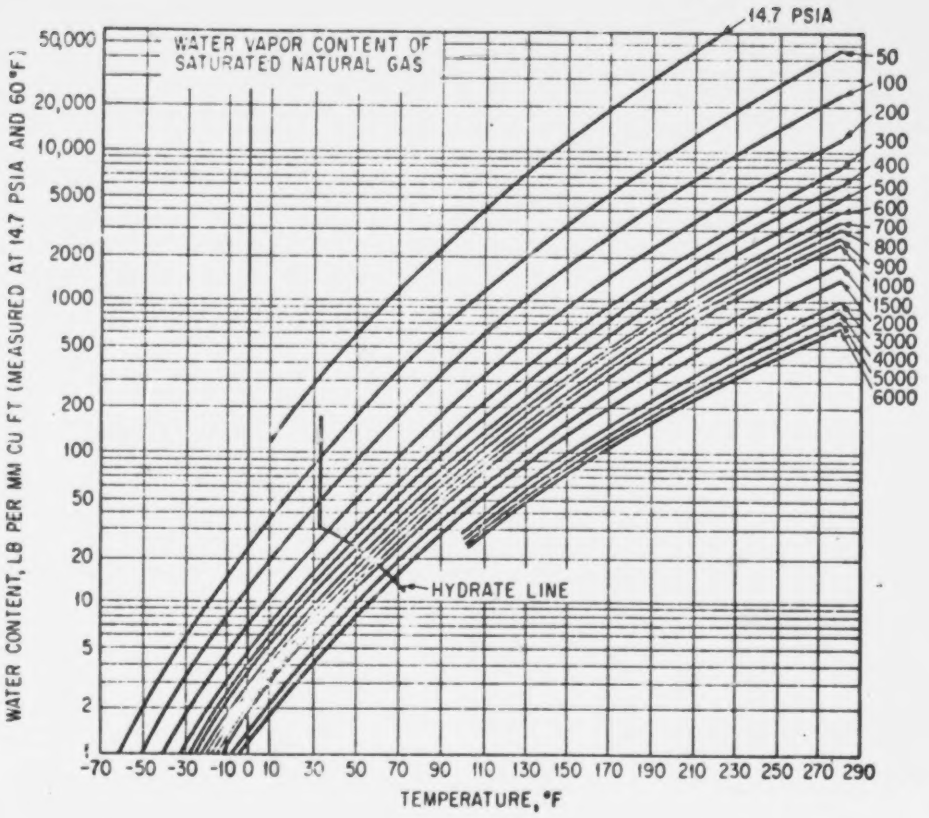


Figure 2. 1. 1. -1. WATER-VAPOR CONTENT OF SATURATED NATURAL GAS[47]

Although it is very light and porous, it can clog small lines. It is thought that hydrates are mechanical mixtures that form when water molecules trap gas molecules (methane and propane) in the water molecule interstices. The unusual facet of this behavior, as can be seen from Figure 2.1.1. -1, is that this solidification can occur well above 32°F, the normal freezing point of water.

To avoid these problems and those related to actual ice formation in the cold box of an LNG plant, it has been determined that water concentration in the gas should be reduced to 1 ppm.[62] The significance of this infinitesimal portion of water in natural gas can be appreciated by referring once again to Figure 2.1.1.-1. One ppm of water would be 0.0416 lb of water/million CF gas, which is off the graph. Most high-pressure natural gas transmission lines carry gas previously dehydrated to a water vapor level of 5-7 lb/million CF.

2.1.1.1.1. Glycol Systems

2.1.1.1.1.1. Glycol System, Process Description

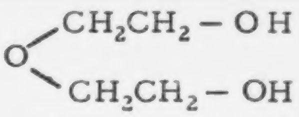
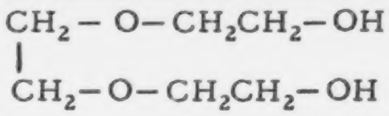
One of the older methods of dehydrating gas is to bring the gas into contact with hygroscopic substances such as diethylene glycol (DEG) and triethylene glycol (TEG). Data on the physical properties of these compounds are presented in Table 2.1.1. -1.

Their unusually high hygroscopic effect, their excellent stability to heat and chemical decomposition, their low vapor pressures, and their ready availability at moderate cost are the factors that have promoted the widespread use of these glycols.

A typical glycol dehydration plant is shown in Figure 2.1.1.-2. A simplified flow diagram of a plant designed to use either of the glycols for the dehydration of natural gas is shown in Figure 2.1.1.1-3. The design shown includes vacuum regeneration and is typical of a large installation where maximum dehydration is required. As can be seen, the process is quite simple. The glycol stream, containing from 1 to 5% water, contacts the gas in a short, countercurrent column. The water, which is absorbed dilutes the glycol somewhat. The dilute solution must be reconcentrated before it can be reused in the absorber. The reconcentration is accomplished by distilling water out of the solution in a regenerator. Because of the extreme difference in boiling points of water and glycol, a very sharp separation can be accomplished with a relatively short column. Some water reflux must be provided at the top of this column to effect rectification of the vapors and minimize glycol losses. In the flow arrangement shown, essentially all of the regenerator vapor stream must be condensed to minimize the load on the vacuum pump (or steam jet), and a portion of this condensate is returned to the column as reflux. With atmospheric-pressure regenerators, it is common practice to condense only the amount of water required for reflux.

The degree of dehydration that can be attained with a glycol solution is primarily dependent upon the extent to which water is removed from the solution in the regenerator. To reduce the water content of the solutions to the absolute minimum without using excessively high temperatures, the process illustrated in Figure 2.1.1.1-3 makes use of vacuum regeneration.

Table 2.1.1.-1. PROPERTIES OF GLYCOLS USED FOR GAS DEHYDRATION[42]

Property	Diethylene Glycol	Triethylene Glycol
Formula	 $O(CH_2CH_2-OH)_2$	 $CH_2-O-CH_2CH_2-OH$ $CH_2-O-CH_2CH_2-OH$
Molecular Weight	106.12	150.17
Specific Gravity (20/20°C)	1.1184	1.1254
Boiling Point (at 760 mm Hg pressure)	473°F (245.0°C)	549.32°F (287.4°C)
Freezing Point	17.6°F (–8°C)	19.04°F (–7.2°C)
Viscosity, absolute (20°C), cp	35.7	47.8
Specific Heat (20°C), Btu/(lb) (°F)	0.500	0.525
Heat of Vaporization (760 mm Hg pressure), Btu/lb	150	179
Heat of Solution of Water in Infinite Amount of Glycol (approx 80°F), Btu/lb	58	86
Initial Decomposition Temp., °F	328	404

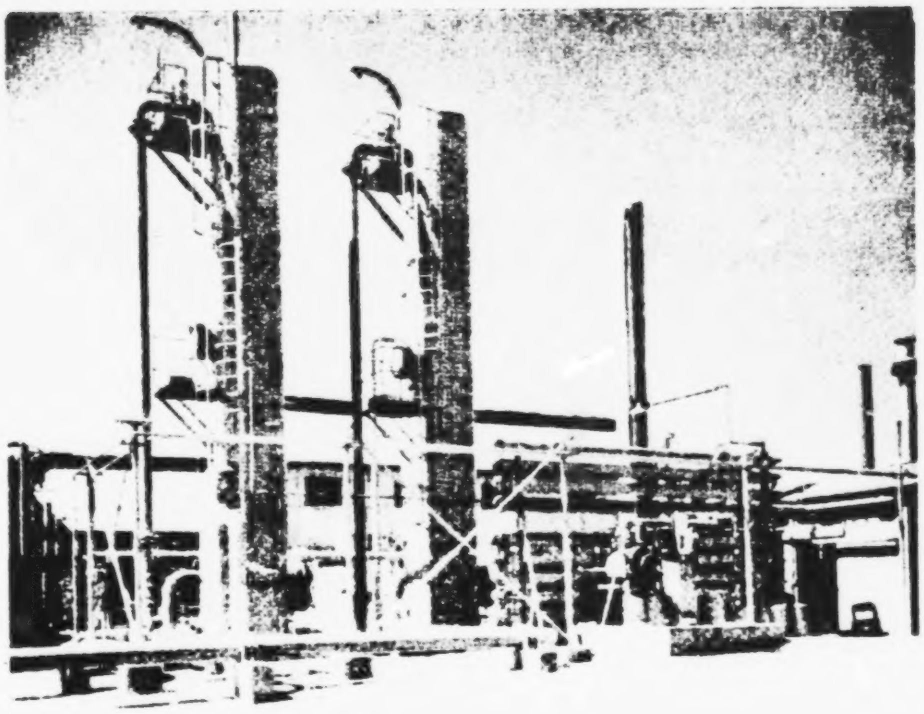


Figure 2.1.1.-2. LARGE NATURAL GAS DEHYDRATION PLANT DESIGNED TO OPERATE AT 1000 psig ABSORPTION PRESSURE[42]

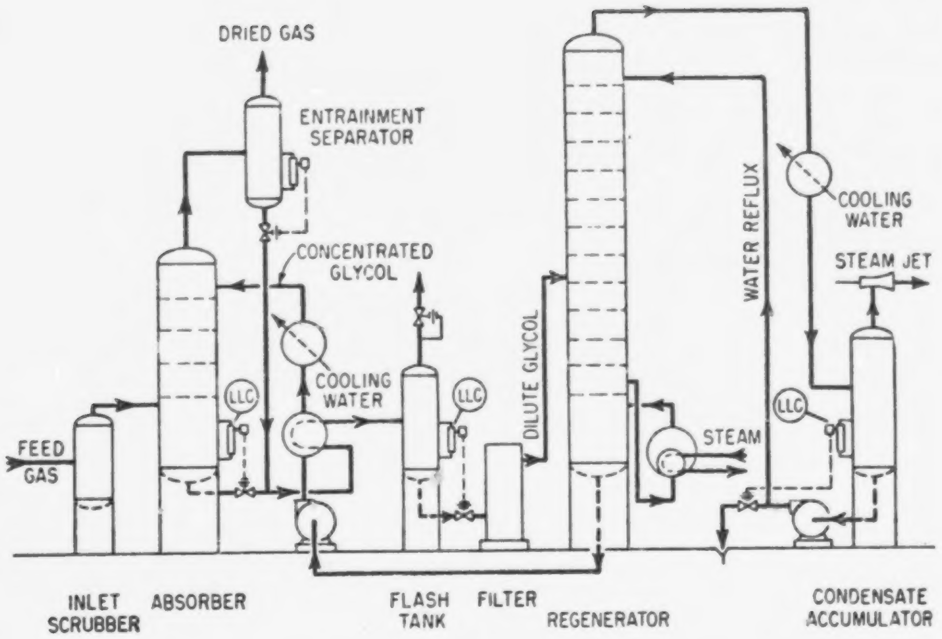

DRIED GAS
ENTRAINMENT SEPARATOR
CONCENTRATED GLYCOL
COOLING WATER
LLC
LLC
FEED GAS
DILUTE GLYCOL
WATER REFLUX
COOLING WATER
STEAM JET
LLC
STEAM
INLET SCRUBBER
ABSORBER
FLASH TANK
FILTER
REGENERATOR
CONDENSATE ACCUMULATOR

Figure 2.1.1-3. FLOW DIAGRAM OF TYPICAL GLYCOL DEHYDRATION PLANT
(Design Shown Would Be Employed Where a Large Volume of High-Pressure Gas Must Be Dehydrated to Extremely Low Water-Vapor Levels)[42]

Calculated dew point depression of dried gas at various circulation rates of TEG solution is shown by Figure 2.1.1.-4.

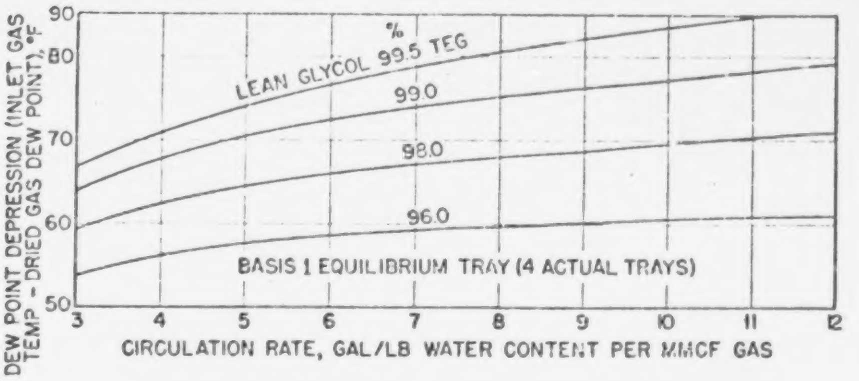


Figure 2.1.1.-4. DEW POINT DEPRESSION OF TEG SOLUTIONS AT VARIOUS CIRCULATION RATES[1]

Dew point depression is the difference between the inlet gas temperature and the dew point of dried gas, both at same contactor pressure, assuming that the incoming gas is saturated. It is apparent that dew point depression varies materially with small changes in the water content of lean glycol solution. If glycol could be regenerated to zero water content, its drying ability would be greatly increased.

A typical dew point depression range is 60°–75°F. Thus, referring again to Figure 2.1.-1 and our example – the 1000-psia gas, saturated at 60°F and containing 19 lb of water/million CF – a dew point depression of 75°F would mean that the gas would be saturated at –15°F and thus would contain approximately 1 lb of water. Therefore, the glycol picks up 1 lb of water. If the TEG glycol solution is 99%, it can be seen that the flow rate required is 8 gal/lb of water content in the dried gas. The regeneration equipment must be sized to boil off 18 lb of water/million CF The TEG will not meet the high standard of 1 ppm for LNG liquefaction. However, it can be used as a pretreatment, the molecular sieve or other method being the final purification step.

2. 1. 1. 1. 1. 2. Investment Costs

The investment cost for TEG dehydration units for field use has been presented as a function of the size of the single unit.[9] These costs have been updated to 1968 by use of the Nelson Cost Index.[54] In general, the investment costs of competitive types of dehydration units indicate that TEG will have the lowest initial investment cost, within the range of water contents for which TEG is suitable.

Capacity, 10^6CF/day	Initial Investment, \$/10^6 CF/day
5	1225
10	925
20	785
30	715
50	655

Costs are based on a 1000-psig contactor and do not include the cost of any auxiliaries required.

If the gas is to be processed at some pressure other than 1000 psig, the cost will be higher. The variation occurs because the contactor size increases rapidly as pressure is decreased, and the contactor vessel weight increases accordingly. Similarly, for higher pressures, the contactor thickness required increases faster than the decreasing volume, which again results in increasing weight.

Contactor Pressure, psig	Cost Correction Factor
400	1.08
700	1.03
1000	1.00
1300	1.05
1600	1.20

2. 1. 1. 1. 1. 3. Operating Costs

The operating costs given below[9] include charges for fuel and chemicals only and do not include operating labor and maintenance costs. These latter costs vary widely from company to company, and depend primarily on the method of cost allocation used. Most recently built dehydration plants are designed for completely automatic operation, with only periodic inspection.

Based on a review of several TEG field dehydration units,[9] the operating cost is estimated to be from $0. 20 to $. 40/million SCF of gas processed.

2. 1. 1. 1. 2. Amine Systems

2.1.1.1.2.1. Amine Systems – Process Description

When glycol systems are combined with amine systems, they are suitable for desulfurization as well as dehydration. The two amines that have proved to be of major commercial interest for gas purification are monoethanolamine (MEA) and diethanolamine (DEA). Triethanolamine has been displaced largely because of its low capacity (resulting from higher equivalent weight), its low reactivity as a tertiary amine, and its poor stability.[42] Figure 2.1.1.-5 shows a relatively large installation for purifying and dehydrating natural gas for pipeline transmission. A combination MEA-DEA glycol solution is used.

Structural formulas for the alkanolamines mentioned above are presented below.

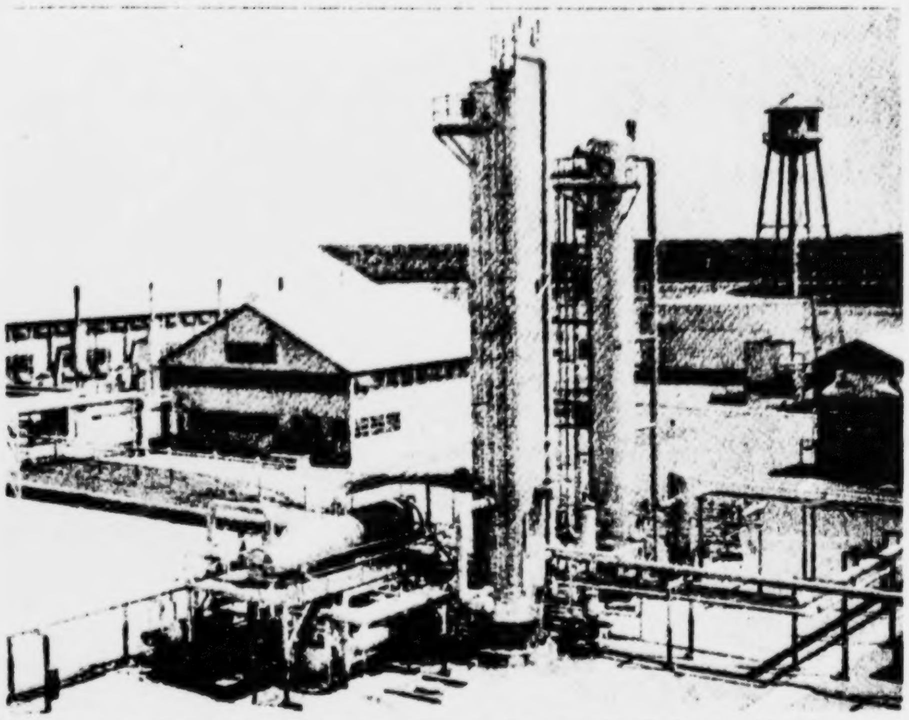
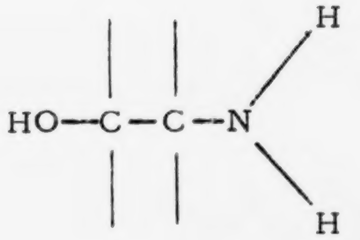
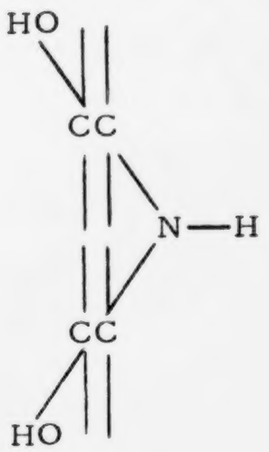


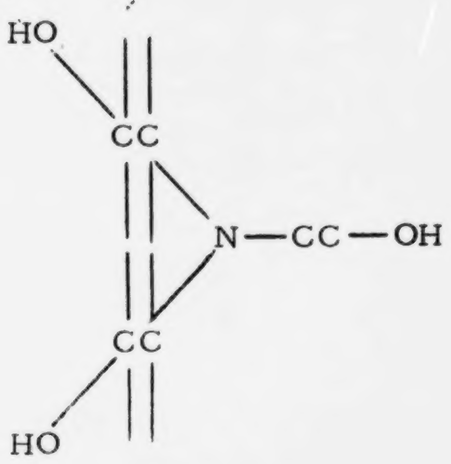
Figure 2.1.1.-5. GLYCOL-AMINE PLANT FOR PURIFYING AND DEHYDRATING HIGH-PRESSURE NATURAL GAS[42]



Monoethanolamine



Diethanolamine



Triethanolamine

Each has at least one hydroxyl group and one amino group. In general, the hydroxyl group serves to reduce the vapor pressure and increase the water solubility, while the amino group provides the necessary alkalinity in water solutions to cause the absorption of acidic gases.

The principal reactions occurring when solutions of a primary amine, such as MEA, are used to absorb CO_2 and H_2S may be represented as follows:

$$\begin{array}{rcll}
2RNH_2 + H_2S & \rightleftharpoons & (RNH_3)_2S & \qquad (2\text{-}1) \\
(RNH_3)_2S + H_2S & \rightleftharpoons & 2RNH_3HS & \qquad (2\text{-}2) \\
2RNH_2 + CO_2 + H_2O & \rightleftharpoons & (RNH_3)_2CO_3 & \qquad (2\text{-}3) \\
(RNH_3)_2CO_3 + CO_2 + H_2O & \rightleftharpoons & 2RNH_2HCO_3 & \qquad (2\text{-}4) \\
2RNH_2 + CO_2 & \rightleftharpoons & RNHCOONH_3R & \qquad (2\text{-}5)
\end{array}$$

Equations 2-3 and 2-4 require water for completion. This water is provided by the original amine solution as well as the vapor in the natural gas. However, it is the glycol wash for amine recovery that is the primary factor in dehydration. The following description indicates the general principles.[42]

The flow pattern for all alkanolamine acid gas absorption process systems is shown in Figure 2.1.1.-6. Gas to be purified is passed upward through the absorber, countercurrent to a stream of the solution. The rich solution from the bottom of the absorber is heated by heat exchange with lean solution from the bottom of the stripping column and then fed to the stripping column at some point near the top. The lean solution, after partial cooling in the heat exchanger, is further cooled by exchange with water or air and fed into the absorber to complete the cycle. Acid gas that is removed from the solution in the stripping column is cooled to condense a major portion of the water vapor. This condensate (or other pure water source) is continually fed back to the system to prevent the amine solution from becoming progressively more concentrated. Generally, all this water, or a portion of it, is fed back to the top of the stripping column at a point above the rich-solution feed and serves to force back amine vapors carried by the acid-gas stream.

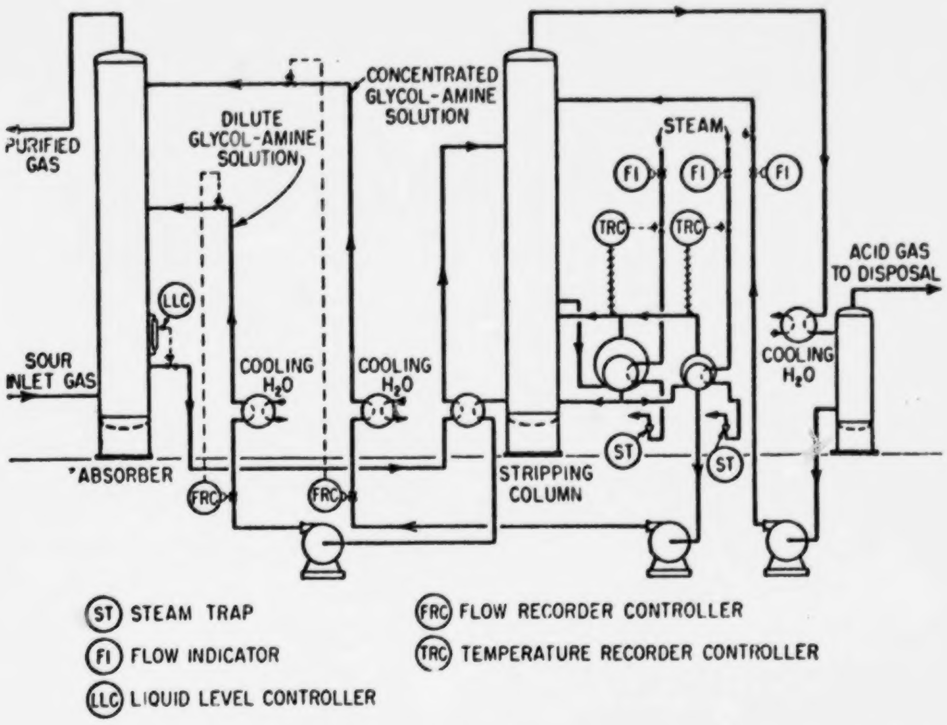

PURIFIED GAS
DILUTE GLYCOL-AMINE SOLUTION
CONCENTRATED GLYCOL-AMINE SOLUTION
STEAM
FI
FI
FI
TRC
TRC
ACID GAS TO DISPOSAL
LLC
SOUR INLET GAS
COOLING H₂O
COOLING H₂O
COOLING H₂O
ST
ST
ABSORBER
FRC
FRC
STRIPPING COLUMN
ST STEAM TRAP
FI FLOW INDICATOR
LLC LIQUID LEVEL CONTROLLER
FRC FLOW RECORDER CONTROLLER
TRC TEMPERATURE RECORDER CONTROLLER

Figure 2.1.1.-6. GLYCOL-AMINE PROCESS MODIFICATION TO MINIMIZE AMINE LOSSES[42]

Obviously, a water wash cannot be used with the glycol-amine system to recover amine vapors from the purified gas stream if dehydration is desired. In Figure 2.1.1.-6a small stream of solution that is relatively low in amine content is used to wash the gas stream after it leaves the primary absorber.

This solution is stripped in a separate small reboiler, the vapors from which may be vented directly into the main stripping column as shown. In one modification, the primary contacting is done at a low pressure to reduce the hydrogen sulfide and carbon dioxide contents of the gas flowing to the compressors. The final wash with a solution that has a low amine and water content serves to remove the last traces of H_2S and CO_2 and provide maximum dehydration with negligible loss of the amine by vaporization. In a commercial operation using an arrangement of this type, the primary solution contains 30% MEA and 13% water, with the balance glycol, while the wash solution contains about 10% MEA and 8% water.

2. 1. 1. 1. 2. 2. Investment and Operating Costs

The use of glycol-amine solutions for dehydration alone is not economic. The glycol in the solution is the only active dehydration component. Therefore, if a 50% amine-50% aqueous glycol solution is used, all of the regeneration system will have to be twice as large to handle the doubled solution circulation rate. This leads to increased investment and operating costs over the glycol system which are not warranted by any increase in performance.

Investment and operating costs for the glycol/amine system, when used for simultaneous acid gas removal and dehydration, are given in Section 2. 1. 1. 2. 5.

2. 1. 1. 1. 3. Molecular Sieves

2.1.1.1.3.1. Molecular Sieve – Process Description

To the geologists, molecular sieves fall into the class of minerals known as zeolites – from the Greek "zeo" and "lithos," meaning "boiling stones." To the chemist, molecular sieves are classified as alkali metal aluminosilicate crystals that can be synthesized in hydrothermal solution. They are classified as zeolites, or "boiling stones," because, unlike other hydrated aluminosilicates and most other hydrated minerals, they can give up water of hydration (e.g., the adsorbent is activated by heating to drive off water) without simultaneous collapse of the mineral structure. The remaining crystal lattice is capable of readsorbing water or other molecules. The strength and integrity inherent to crystalline materials enables molecular sieves to withstand repeated adsorption-desorption cycles almost indefinitely.

The atoms of oxygen, silicon, aluminum, and alkali metal — usually sodium, potassium or calcium — which make up molecular sieves are all tightly bound into the crystal structure in a very uniform three-dimensional array. Figure 2.1.1.-7 depicts a unit cell of molecular sieve Type 4A, with each white sphere representing a single oxygen atom and each dark sphere a single sodium atom. The silicon and aluminum atoms are not shown because they are relatively small and their presence is not essential to this part of the discussion.

It can be readily seen that the unit cell is basically a cube with an aperture in each of its six faces. Each aperture is formed by exactly eight oxygen atoms and leads directly into the interior cavity of the cell. In the case of molecular sieve Type 4A, the diameter of each aperture is about 4 angstroms, or about 16 billionths of an inch, and the diameter of the cavity is approximately 11 angstroms. The aperture diameters approximate the minimum effective diameter of certain individual molecules that can enter the internal cavities from which larger molecules are excluded, hence the name "molecular sieves."

By controlling certain variables — such as the ratio of ingredients, the alkali metal cation, and the conditions of the hydrothermal solutions in which these crystals are literally grown during their manufacture — it is possible to synthesize other configurations with different aperture diameters.

A full-grown crystal is comprised of a three-dimensional array of unit cells so arranged that the internal cavities are interconnected by the uniform apertures. Any molecule small enough to enter an aperture is able to migrate freely through the maze of internal cavities. A single typical crystal measuring approximately 1.25 microns (about 50 millionths of an inch) on each edge contains about 1 billion unit cells.

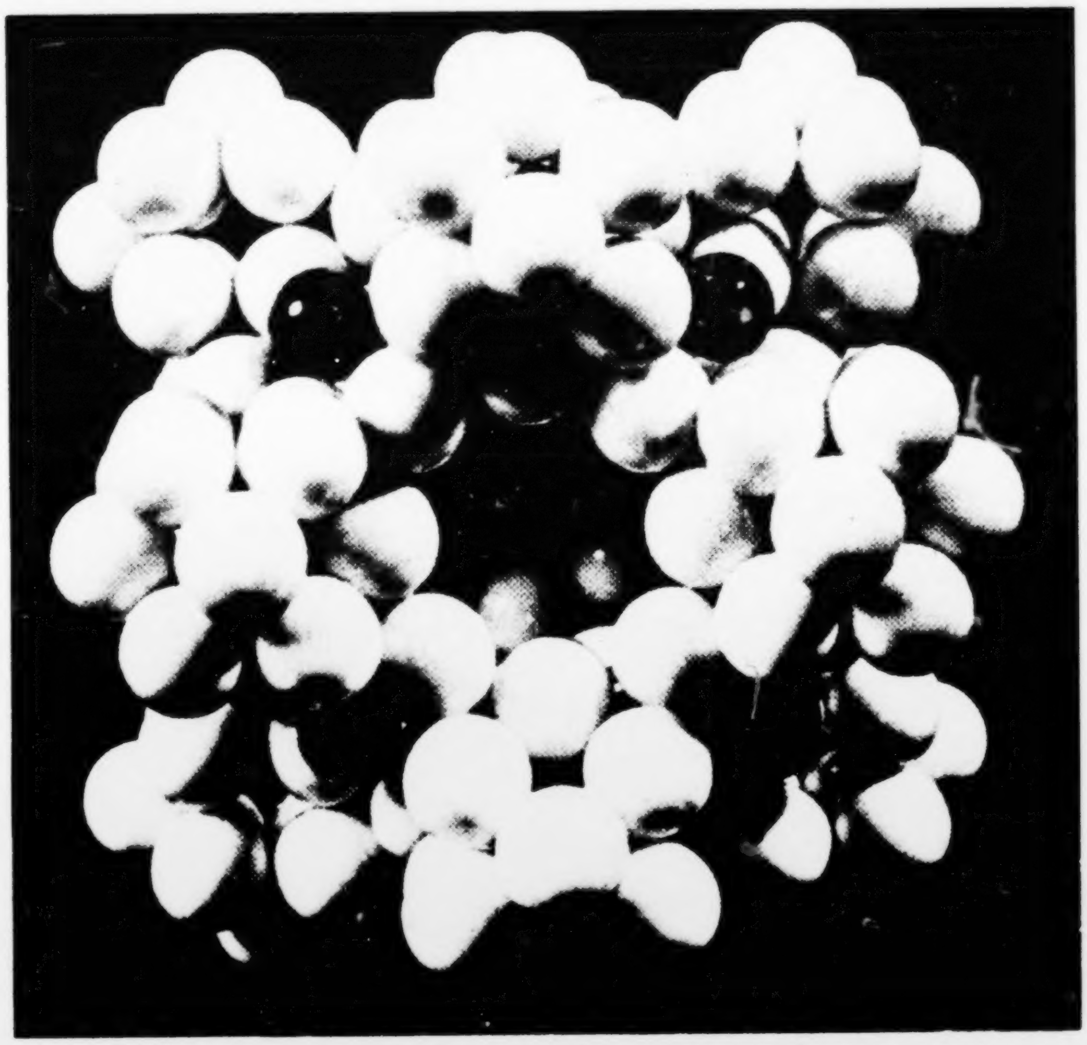


Figure 2.1.1.-7. UNIT CELL OF MOLECULAR SIEVE TYPE 4A (White Spheres Represent Oxygen Atoms; Dark Spheres Represent Sodium Atoms)[62]

To make these crystals useful, an inert binder is added and the resultant mixture is formed into pellets that typically measure 1/16 or 1/8 in. in diameter. The surface area of molecular sieve pellets is about 500-750 sq m/g, with the surface of the internal cavities comprising approximately 99% of the total surface area.

Some of the outstanding properties of molecular sieves are readily illustrated by comparison with such other common adsorbents as silica gel and activated alumina.

1. Molecular sieves are crystalline. The other common adsorbents are amorphous; i.e., they have no repetitive structure and can be described only in a statistical sense.
2. Because they are crystalline, molecular sieves possess excellent physical properties. Amorphous desiccants tend to spall or shatter when contacted with water, which causes thermal stresses to be created by heat of adsorption. Molecular sieves do not spall, nor do they tend to cake if contacted with caustic.
3. The alkali metal cations in the molecular sieve structure generate an electrostatic field within each internal cavity, causing the surfaces to be polar. Accordingly, molecular sieves possess very high affinity for polar molecules. In contrast, adsorption on the amorphous adsorbents involves relatively weak surface attraction forces.
4. Table 2.1.1-2 illustrates how the uniform aperture size of molecular sieves permits selective adsorption. For example, Type 3A is specific to water and ammonia only; all hydrocarbons are excluded from the internal cavities. This molecular sieving property has been used very effectively in natural gas dehydrators, for example, to eliminate coadsorption and, thereby, to provide uniform high capacity and long service life by eliminating polymerization and other fouling of the active sites.
5. Typical isotherms (Figure 2.1.1-8 show that molecular sieves possess high capacity at temperatures as high as 200° F. Silica gel and alumina, on the other hand, are very sensitive to minor temperature changes at typical processing conditions and retain little, if any, adsorptive capacity at 100° F.
6. The typical isobars presented in Figure 2.1.1-9 show that molecular sieves possess uniform high water capacity even at very low relative humidity or low partial pressure and are thus capable of drying to extremely low dew points.

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Table 2.1.1.-2. MOLECULAR SIEVES PERMIT SELECTIVE ADSORPTION[62]

BASIC TYPE	NOMINAL PORE SIZE (ANGSTROMS)	AVAILABLE FORM	BULK DENSITY (LB/CU FT)	EQUILIBRIUM H_2O CAPACITY (WT %)*	MOLECULES ADSORBED**	MOLECULES EXCLUDED	OUTSTANDING FEATURE
3A	3.0	Powder 1/16" Pellets 1/8" Pellets	30 47 47	23 20 20	H_2O NH_3	Ethane and larger	Dries polar liquids and gases such as methanol, ethanol, propylene, butene and cracked gas.
4A	4.0	Powder 1/16" Pellets 1/8" Pellets 8 x 12 Beads 4 x 8 Beads	30 45 45 45 45	28.5 22 22 22 22	Ethanol, H_2S, CO_2, SO_2, C_2H_6, C_2H_4, C_3H_6	Propane and larger	Dries non-polar liquids and gases, such as natural gas, to minimum effluent water content with maximum capacity.
5A	5.0	Powder 1/16" Pellets 1/8" Pellets	30 43 43	28 21.5 21.5	n-C_4H_9OH*** n-C_4H_{10}*** C_3H_8 to $C_{22}H_{46}$ R-12	Iso Compounds and all 4 carbon rings and larger	Separates normal from iso hydro-carbons.
13X	10	Powder 1/16" Pellets 1/8" Pellets	30 40 40	36 28.5 28.5	Di-n-propylamine Molecules < 9Å	$(C_4F_9)_3N$ and larger	Coadsorbs water and H_2S or CO_2

* Lb H_2O/100 lb activated adsorbent at 17.5 mm Hg, 25 °C. ** Each type adsorbs listed molecules plus those of preceding type. *** Also larger family members.

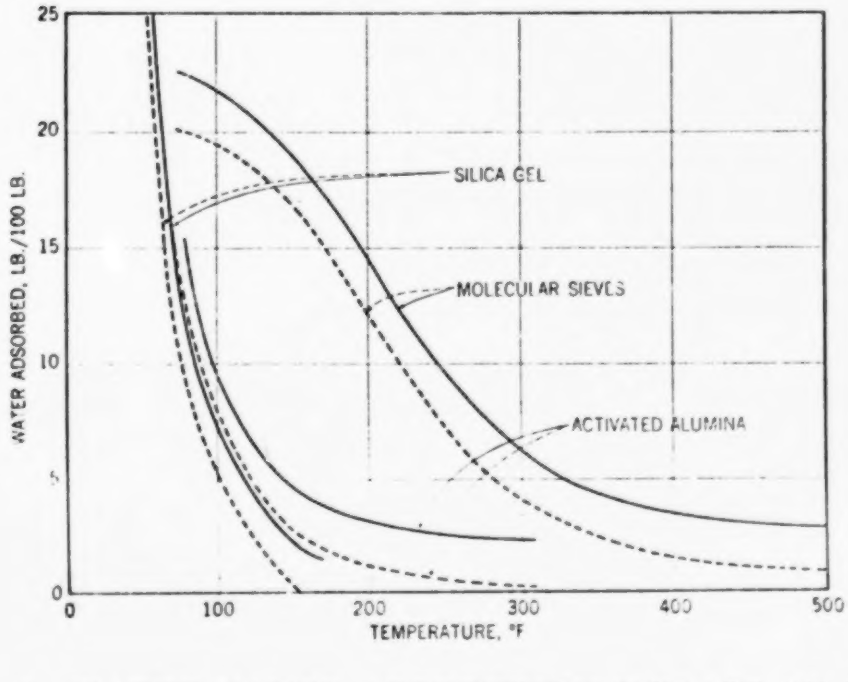

Figure 2.1.1.-8. WATER VAPOR ADSORPTION ISOBARS AT 10 mm PARTIAL PRESSURE (Equilibrium Data)[62]

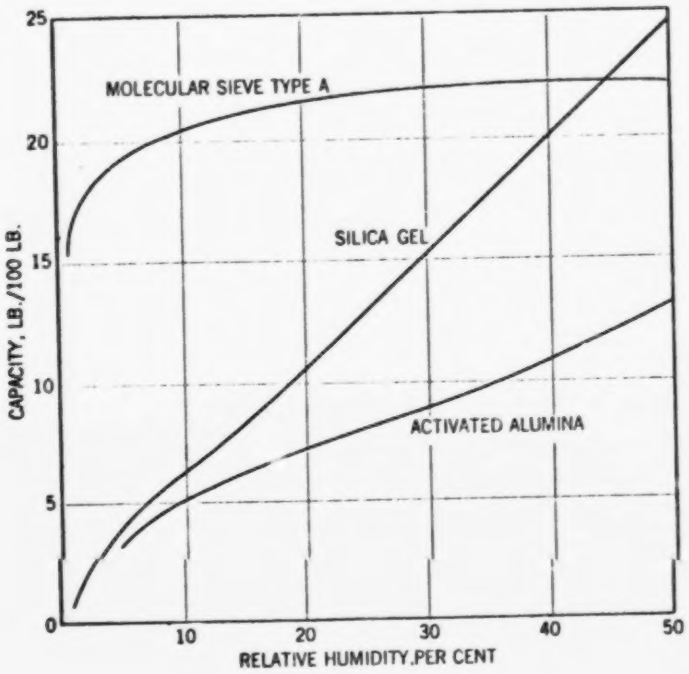

Figure 2.1.1.-9. WATER VAPOR ADSORPTION AT 25°C (Equilibrium Data)[62]

As an example, the experience of San Diego Gas & Electric with its molecular sieve treatment is that approximately equal quantities of oil and water are being removed. All evidence indicates an oil-free cold box in its LNG plant. This oil is trapped on the surface of the pellet rather than by cationic attraction. A molecular sieve natural gas prepurification process consists of essentially the same equipment, and involves about the same operation, as an ordinary solid-bed gas dehydration system. A typical flow diagram for a dual-bed unit is shown in Figure 2.1.1.-10. While one bed is adsorbing, the second bed is regenerating. The adsorber beds are usually designed for relatively short cycle times. The bed configuration is such that the pressure drop during adsorption is usually of the order of only 1-2 psi.

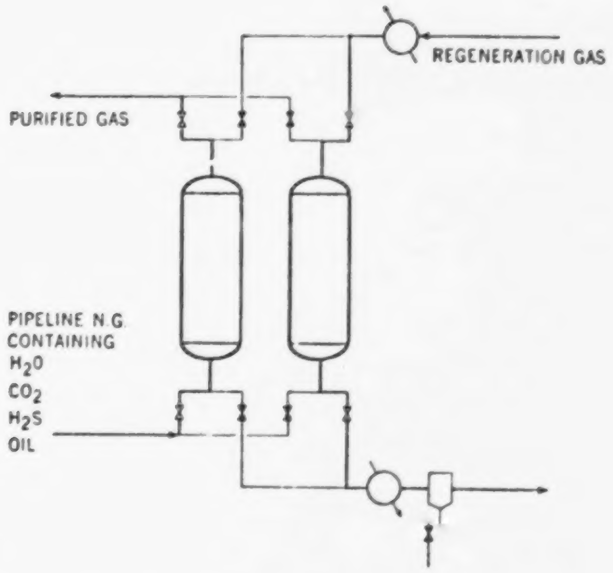

Figure 2.1.1.-10. TYPICAL FLOW DIAGRAM FOR A DUAL-FLOW UNIT[62]

Depending upon such factors as the particular cycle and the location of the liquefaction plant, regeneration may be effected at essentially the same pressure as adsorption (e.g., pipeline pressure) or at some lower pressure (e.g., the distribution system pressure). Regeneration requires a heater, cooler, and knockout drum as the only auxiliary equipment. Automatic control valves are used. The molecular sieve adsorbers are designed for unattended operation.

Adsorption consists of taking gas from the pipeline, passing it through a mechanical entrainment separator, and then directing it through the on-stream adsorber and on into the cold box header. Immediately after switching from adsorption to regeneration (a step that might also involve depressuring), regeneration gas passes through the heater and on into the bed to desorb the contaminants. Exit gas passes through the cooler-knockout combination for recovery of water and oil and then back into the pipeline.

At San Diego Gas & Electric, regeneration gas flows directly to a nearby power plant. Note that in this arrangement the regeneration gas carries the desorbed carbon dioxide back into the pipeline, so there are no acid gas disposal problems. Equally important, there is no volume shrinkage. After sufficient heating and desorption, the inlet gas heater is disconnected and the bed is cooled. At completion of regeneration, the bed is switched back to adsorption. Of course, the valve change may also involve repressuring.

Now, the logical question is: "Where does the regeneration gas come from?" Boil-off and tail gas are two possible sources, but the volume may not be sufficient or they may not be available during initial start-up. Domestic LNG facilities will be located along pipelines or at a distribution depressuring point. Typically, only a fraction of the total gas passing through or past the liquefaction plant will actually be liquefied. Even at "total liquefaction plants," such as those considered for North Africa, a significant fraction of the total gas stream is consumed for onsite energy requirements. Obviously, then, a small portion of the pipeline gas that would otherwise bypass the LNG plant or the gas consumed for onsite energy can be used for regeneration. The regeneration gas requirement drawn from these sources is purified along with the gas to the cold box and thereby becomes an excellent clean regeneration media.

2.1.1.1.3.2. Investment Costs

The investment cost for a relatively large gas dehydration unit for the feed to a low-temperature natural gasoline plant has been reported, and is summarized below.[69]

Feed:	1 X 10^9 SCF/day of natural gas
Pressure:	600 psig
Temperature:	80°F
Water Content,	
in:	12 lb water/10^6 SCF
out:	< 0.1 ppm water
Dehydrator:	4 beds — 12 ft diam by 12 ft deep
Molecular Sieve:	250,000-lb Type 4A
Pressure Drop Through Contactor:	7-1/2 psi
Fuel:	5 million Btu/hr
Investment:	$640,000 installed, including cost of sieves

2.1.1.1.3.3. Operating Costs

The use of molecular sieves requires slightly more energy for regeneration than for other solid dessicants. However, the extended life and capacity of the sieves and reduced shutdowns due to water vapor carry-over more than offset the increased energy required. The commercial acceptance of molecular sieves can be illustrated by the recent change-over of five natural gas treating plants at helium recovery facilities. Reasons for the change are given below.[69]

Plant	Water Content Entering, lb/10^6 SCF	Previous Method	Reason for Change to Molecular Sieves
A	40	Alumina plus sieves	Poor alumina life
B	40	Alumina plus sieves	Poor alumina life
C	100	Glycol	Freeze-ups
D	3-5	Alumina	Poor alumina life plus freeze-ups plus high ΔP
E	3-5	Glycol	Freeze-ups

Operating costs have been reported as $0.60-$1.20/million CF processed.[9]

2.1.1.1.4. Other Solid Adsorbents

A large percentage of the gas dehydrated in the U.S. is processed by dry desiccant units. Development of liquid desiccant systems has been rapid. Under general field conditions these newer liquid systems have an economic advantage over the dry desiccant units. However, dry desiccants have retained their position in many applications for which they are uniquely qualified: achieving very low dew points in the gas stream, especially prior to liquefaction, and simultaneous dehydration and recovery of liquid hydrocarbon components.

2.1.1.1.4.1. Process Descriptions

Solid desiccants occur in two general classes. The most effective drying agents are those that react chemically with the water. These include magnesium perchlorate, calcium oxide, barium oxide, and anhydrous calcium sulfate. The problem is that those materials tend to form such a strong bond that regeneration is very expensive.

The second type of desiccants owe their activity to surface adsorption and capillary condensation. The physical properties of some of these adsorbents are given in Table 2.1.1.-3. These are the desiccants important to dehydration, since they can be regenerated quite easily. However, this ease of regeneration means they are not quite as effective drying agents, although they can easily give dew points well below those normally required for gas pipeline operation. The relative drying efficiency of six common desiccants is given below.[12]

Material	Dew Point, °F
Silica Gel	−62
Sovabeads	−75
Activated Alumina	−87
Activated Bauxite (Florite)	−87
Magnesium Perchlorate	−96
Calcium Oxide	−95

The temperature at which the moisture is adsorbed has a pronounced effect on the water capacity of some desiccants, but only a minor effect on others, as shown in Figure 2.1.1.-11. Also, the velocity of the gas flowing through the bed has an appreciable effect on capacity, as shown in Figure 2.1.1.-12. Although the various desiccants vary in the dew point they can achieve, all give gas with much lower dew point than required to meet pipeline specifications.

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Table 2.1.1.-3. PROPERTIES OF THE ADSORPTION-TYPE DESICCANTS[12]

Type of Desiccant	Bulk Density, lb/ft	Specific Heat, Btu/lb/°F	Thermal Conductivity, Btu/hr/ft-°F/in.	Type	Normal Mesh Sizes Used	True Specific Gravity	Percent Voids as Packed
Activated Alumina	51	0.24	1.0	Hard granules	1/2-1/4 in. 1/2 in.-8 mesh 8/14	3.9	49*
S/V Sovabead	48	0.25	1.0	Hard smooth gel	4/8	2.3	40
Florite	50	0.24	1.0	Hard granules	2/4 4/8 4/14	3.4	40†
Alumina Gel	45	0.24	--	Hard spheres	1/4 in.	--	46
Silica Gel	39	0.22	1.0	Hard smooth gel	4/8	2.2	50

* 1/4 in.-8 mesh.

† 4-8 mesh.

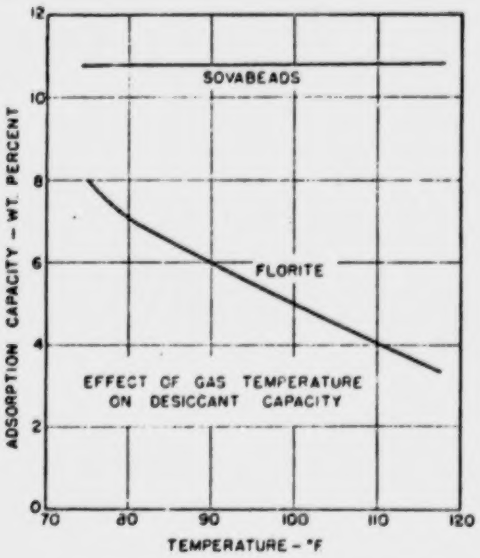


Figure 2.1.1.-11. EFFECT OF GAS TEMPERATURE ON DESICCANT CAPACITY[12]

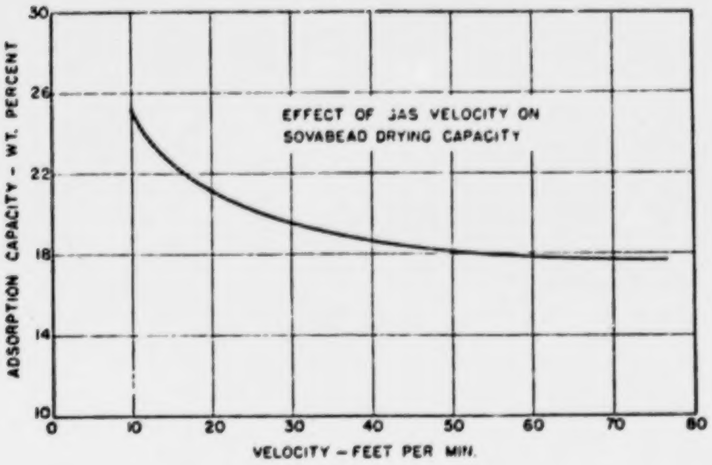


Figure 2.1.1.-12. EFFECT OF GAS VELOCITY ON SOVABEAD DESICCANT CAPACITY [12]

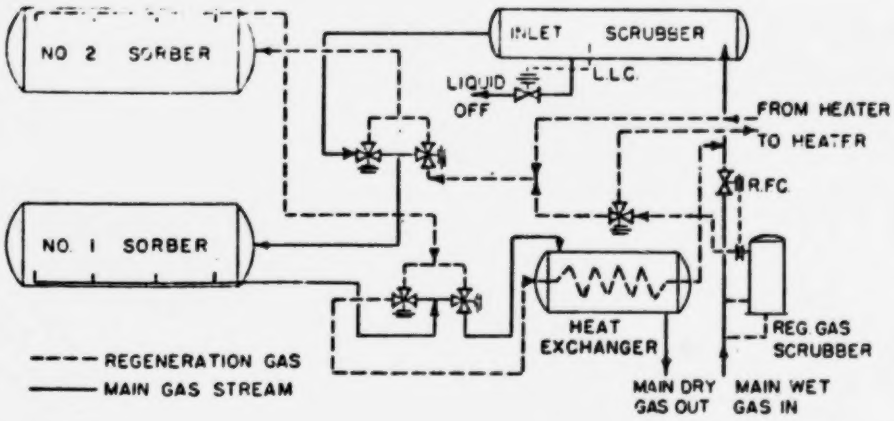

NO 2 SORBER
INLET SCRUBBER
LIQUID OFF
L.L.C.
FROM HEATER
TO HEATER
R.F.C.
NO. 1 SORBER
HEAT EXCHANGER
REG GAS SCRUBBER
REGENERATION GAS
MAIN GAS STREAM
MAIN DRY GAS OUT
MAIN WET GAS IN

Figure 2. 1. 1.-13. FLOW SHEET OF NATURAL GAS DESICCANT DEHYDRATOR [12]

The flow diagram for a typical two-column dry desiccant dehydration unit is shown in Figure 2. 1. 1.-13. The wet gas from the separator splits into two streams, one for dehydration (usually 85 to 95% of total flow) and one for regeneration. The main gas stream flows through an inlet scrubber where any liquids, both water and hydrocarbons, are removed. The scrubbed gas enters a cyclic flow control valve and is directed to the adsorber presently in use for dehydration (in Figure 2. 1. 1.-13, the No. 1 sorber). The dried gas flows through a second set of cyclic flow control valves to a heat exchanger, where it is used to cool the wet regeneration gas, and from there to the pipeline.

The regeneration gas passes through a small gas scrubber to a cyclic flow control valve. During the regeneration heating cycle, this gas is sent to a heater where its temperature is raised to 400°-450°F. The hot gas from the heater enters the cyclic flow control valves and is directed to the vessel being regenerated (in Figure 2. 1. 1.-13, the No. 2 sorber). The water-wet gas from the sorber flows through the second set of cyclic valves to the heat exchanger. Here it is cooled by heat exchange with the main dry-gas stream and rejoins the main gas stream prior to the liquid scrubber.

To operate, this system needs a pressure drop at the recording flow controller in the main gas stream. The required pressure drop is that necessary to cause the regeneration gas to flow through the various cycle components and return to the major stream. Some of the important operating characteristics and regeneration practices are given in Tables 2.1.1.-4 and 2.1.1.-5.

Table 2.1.1.-4. SUMMARY OF THE OPERATING CHARACTERISTICS OF DRY DESICCANT DEHYDRATION[12]

Recommended Operating Range: At temperatures preferable below 125°F to utilize the higher desiccant capacity. The gas should enter the unit above its hydrate point. There seem to be no effective pressure limitations.

Desiccant Service: Normally from 1-3 years in the absence of poisoning. Service limited by loss of capacity, dusting, and breakage.

Dehydration Obtained: Essentially bone-dry gas produced (less tha 1 lb water/10^6 SCF). Most units will give a −30°F dewpoint during the first part of the cycle.

Length of Cycle: 8-hr cycle most common for convenience, although others sometimes used 6, 12, and 24 hr.

Regeneration: Final regeneration temperature varies with type of desiccant, usually 350° - 450°F. About 5 to 15% of the total gas stream is used for regeneration. Normal 8-hr cycle uses about 6 hr for heating and 2 hr for cooling.

Advantages:

1. Low exit dew point obtained.
2. Gives effective dew point depressions over a wide range of operating conditions.
3. Compact, particularly the smaller units.
4. Relatively low initial investment for small amounts of gas, such as for instruments where "batch" or "semi-batch" operation, is satisfactory.
5. Rated capacity may be increased if some wet gas is bypassed around the unit and recombined with dry gas, where the unit gives an exit dew point lower than that needed or specified.
6. Some operators report additional light hydrocarbon recovery by adsorption on the desiccant.

Table 2. 1. 1. -4, Cont. SUMMARY OF THE OPERATING CHARACTERISTICS OF DRY DESICCANT DEHYDRATION [12]

Disadvantages:

1. High initial investment.
2. Desiccant sensitive to poisoning, particularly from heavy lube oils, which frequently requires expensive desiccant replacement.
3. Rated capacity of the unit declines with pressure.
4. Pressure drop is higher than with liquid desiccant systems. (In compressed gas systems this substantially increases the required compressor horsepower).
5. At flow rates below the rated capacity, the regeneration heat load is high in relation to the amount of gas processed.

Table 2. 1. 1. -5. SUMMARY OF REGENERATION PRACTICES[12]

Method of Heating: Usually the part of the main stream that is bypassed is either returned to the system downstream or is recombined and dehydrated. Liquefied petroleum gases that have been vaporized or superheated steam are also used.

Required Temperature: In most instances 350°-450°F is recommended. Higher temperatures increase capacity, but cause a shorter effective desiccant life. Majority of water is driven off at 250°F. Gas temperature from heater usually 400° and 450°F.

Direction of Gas Flow: Usually countercurrent to main flow in vertical towers. Other designs often use concurrent flow.

Gas Flow Rate: About 5-15% of main gas stream. Percentage varies with main gas flow rate and with design of unit. Flow rate must be sufficient to supply the necessary heat, in time allotted, for regeneration.

Water Removal: Where regeneration gas recombined with main gas, water from regeneration is removed by cooling with main gas stream or by water cooler. Maximum heat load occurs when bed reaches about 250°F, at which time most of water is coming off. Heat exchanger should be designed for this condition.

Heat Required: Heat of desorption of water plus sensible heat to bring water vapor, desiccant, and container up to temperature. Latent heat of water is often used for heat of desorption in design.

Time Required: To bring exit regeneration gas up to desired temperature. Usually about 65-75% of total cycle time.

Pressure: Full line pressure satisfactory.

2. 1. 1. 1. 4. 2. Investment Costs

The initial cost of dry desiccant plants varies with gas conditions and field site. However, data on a series of field units have been prepared[12] and updated to 1968 prices.[54] These values are based on 1000 psig contactor pressure at 80° F with a 5% desiccant capacity for moisture (by weight). Below is given the investment costs for various daily dehumidification rates.

Dehumidification Rate, 10^6 SCF / day	Investment Cost, \$$10^6$ SCF/day
5	3850
10	2850
20	2070
35	1650
50	1570

Based on this schedule, a dry desiccant plant to dehumidify 20 million CF/day at 1000 psig would cost $41, 700. The effect of pressure on investment is shown in Figure 2. 1. 1. -14.

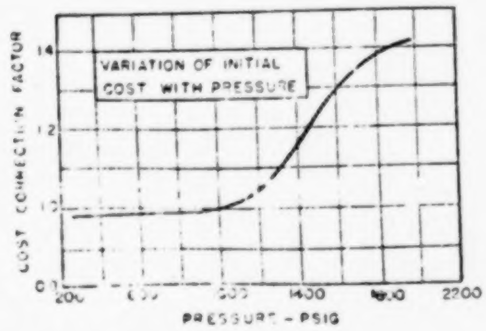


Figure 2.1.1. -14. EFFECT OF GAS PRESSURE ON DRY DESICCANT DEHYDRATOR COSTS[12]

2.1.1.1.4.3. Operating Costs

Operating costs necessarily vary among the many plants. However, most reports[12] show costs ranging from \$0.60 to \$1.20/million SCF. This variation is due primarily to labor changes and special operating problems.

2.1.1.2. Hydrogen Sulfide/Carbon Dioxide Removal

2.1.1.2.1. Caustic Treating

2.1.1.2.1.1. Process Descriptions

The caustic-treating, or "lime," process is one of the earliest methods for removing acid gases from gas mixtures. This is a once-through process with a nonregenerable product and, therefore, is only suitable for handling small volumes of gas with low concentrations of the impurities to be removed. It has not been applied to natural gas.

The caustic-treating process is not considered either practical or economical for natural gas because of the cost of the caustic and the problem of waste product disposal.

2.1.1.2.2. Iron Oxide Dry Box Treating

2.1.1.2.2.1. Process Description

A schematic flow diagram of a typical iron oxide dry box system is shown in Figure 2.1.1.-15.

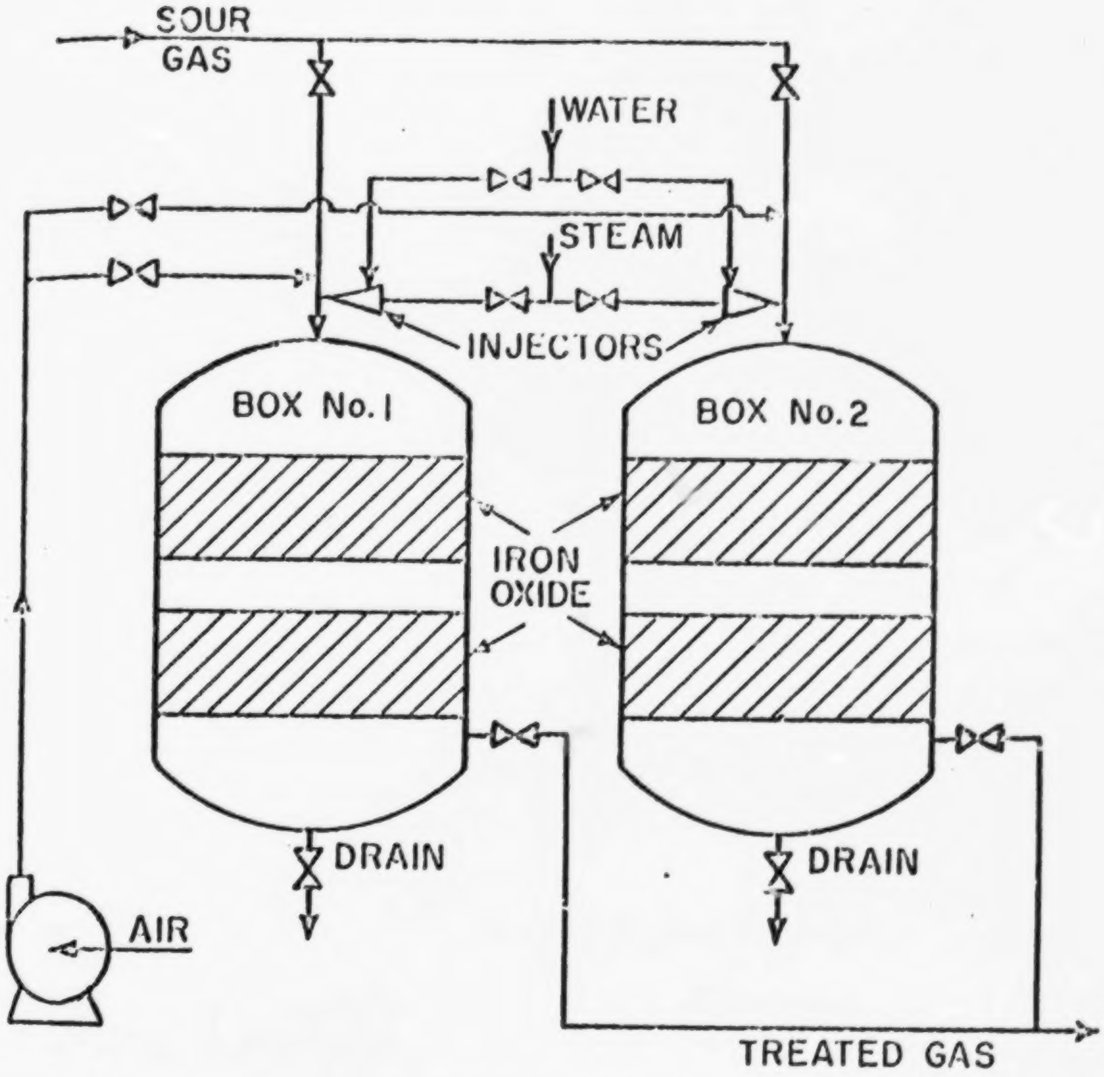

SOUR GAS
WATER
STEAM
INJECTORS
BOX No. 1
BOX No. 2
IRON OXIDE
DRAIN
DRAIN
AIR
TREATED GAS

Figure 2. 1. 1. -15. SCHEMATIC FLOW DIAGRAM OF IRON OXIDE DRY BOX PROCESS

At least two reactors are required for continuous operation because the process is cyclic. Many forms of iron oxide are used; the commonest, iron sponge, consists of iron-oxide-impregnated wood shavings. Iron sponge usually contains from 5 to 10 pounds of Fe_2O_3 per cubic foot of material.

The dry box process operates best at temperatures of 80° to 100°F, water contents of 30 to 50% of the sponge weight, and in the presence of an available alkali, such as Na_2CO_3. The process is completely selective for H_2S in the presence of CO_2 and hydrocarbons, but, if COS, CS_2, or HCN are present, they may be partially removed.

Principal disadvantages of this process are the enormous quantity of material required for treatment of large volumes of gas, the cyclic operation, and the high operating cost. Its chief advantage is its ability to remove traces of H_2S from low-pressure streams containing tars, air, CO_2, HCN, NH_3, COS, or CS_2 which might cause operation difficulties for some of the continuous processes. The product will contain about 0.25 grains of H_2S/100 SCF of treated gas.

2. 1. 1. 2. 2. 2. Investment and Operating Costs

The only application of the dry box process to natural gas processing has been for removal of very small quantities H_2S from a high-CO_2-content gas. Because of the large bulk of the iron-oxide-impregnated support material, the process requires large vessels, which makes the investment cost very high for high-pressure processing. Also, the life of the reacting material is limited, and the accumulation of sulfur in the reacting material created by the regeneration cycles results in a high pressure drop through the reactors. Both effects make the operating costs very high when compared to other H_2S removal processes.

2.1.1.2.3. Molecular Sieves

2.1.1.2.3.1. Process Descriptions

Molecular sieves are crystalline structures grown from alkali metal alumino silicates that have precisely uniform properties. The structures contain pores of uniform size which give the material a very large surface area per unit volume. The surface area within the structure of the molecular sieve is adsorptive toward any molecule that can pass into the pores. However, adsorptive sites within the pores have a greater affinity toward molecules of materials that are polar.

Such polar contaminants include H_2S, CO_2, water, and other sulfur- and oxygen-bearing components. Methane and other hydrocarbons contained in natural gas are nonpolar and therefore almost unaffected by the adsorptive property of the molecular sieves. Pore size for the molecular sieve is also so small that relaively small molecules such as H_2S and water are free to enter, but larger molecules such as CO_2 can be excluded because of effective size. Of course, molecular sieves with larger pore sizes are available and are capable of adsorbing CO_2 and organic sulfur compounds.

Since the particular types of compounds are characterized by varying degrees of adsorptivity, the difference in selectivity of the molecular sieves toward the polar contaminants permits one to design the adsorbers to affect drying of the gas, drying and sulfur removal, or drying and total acid-gas removal (including CO_2).

Several molecular-sieve processes, usually characterized by their regeneration scheme, are available. The sour gas is sweetened by passing it through a bed of molecular sieves. Dehydrated and sweetened gas is recovered. As the activated sites in the sieve pores become filled with adsorbed material, the effectiveness of the bed is reduced. Finally it becomes necessary to regenerate the bed.

One molecular-sieve sweetening operation is reported by Rushton and Hays.[61] Here a 2-tower unit is operated to selectively remove H_2S from a 500-psig gas with 25-20 gr H_2S/100 SCF. The unit can process up to 5 million SCF/day gas to a "no trace" sulfur content.

A schematic flow diagram of the process (Figure 2. 1. 1.-16) shows how tower regeneration is effected. Sour gas is sweetened in Tower 1 while regeneration and cooling takes place in Tower 2. A small portion of the sweetened gas is diverted through the regeneration heater to Tower 2. This hot gas passes through the spent bed and, until the bed reaches a temperature high enough to drive off the adsorbed H_2S, it is returned to the sweetened gas line.

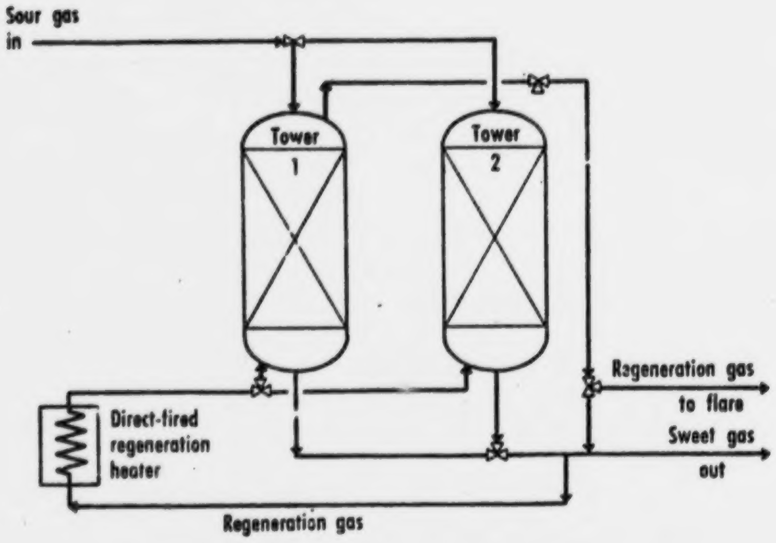


Figure 2. 1. 1.-16. TYPICAL MOLECULAR-SIEVE SWEETENING PROCESS[51]

When the bed reaches the regeneration temperature (about 650°-700°F), this gas is diverted to a flare. The regeneration cycle continues with the heater off until the bed cools to below the driveoff temperature. At this point the regeneration gas is again routed into the sweetened-gas line. The regeneration gas lost to the flare and the direct-fired heater amounts to 1-2% of the daily throughput.

A second process[55] has been developed by Haines and by Krell Associates, Inc., and is shown schematically in Figure 2. 1. 1.-17. Here three towers are employed — one for the adsorption cycle, one for the regeneration cycle, and the third for cooling following regeneration.

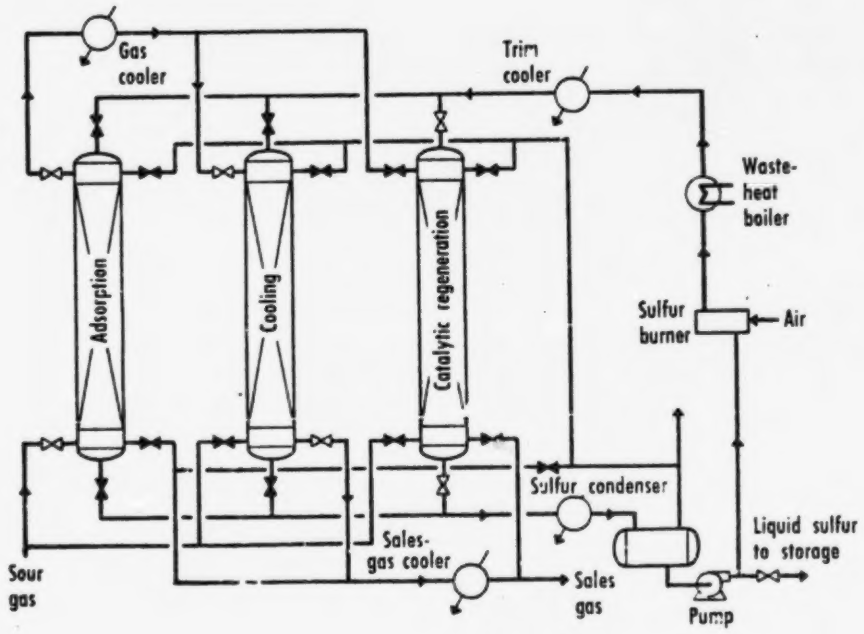

Gas cooler
Trim cooler
Waste-heat boiler
Adsorption
Cooling
Catalytic regeneration
Sulfur burner
Air
Sulfur condenser
Liquid sulfur to storage
Sour gas
Sales-gas cooler
Sales gas
Pump

Figure 2. 1. 1.-17. HAINES PROCESS WITH SULFUR RECOVERY[51]

The process operates on relatively short cycles of about 1-3 hr. The gas flow passes in series through two towers. Sour gas passes through the adsorption tower where it is sweetened, goes through a gas cooler, and then passes through the tower on the cooling cycle to cool the freshly regenerated absorbent. From here it is cooled and then sent to sales.

Regeneration of the third tower is accomplished by stripping the adsorbed H_2S from the molecular-sieve bed with hot sulfur dioxide gases. These gases are generated by burning liquid sulfur produced by the regeneration process itself.

Approximately one-third of the recovered sulfur is burned to produce the sulfur dioxide gases. The gases pass through a waste-heat boiler, a temperature-control cooler, and then through the tower. As the adsorbed H_2S is stripped from the bed, it reacts with the sulfur dioxide to form sulfur vapor and water.

These vapors, some unreacted H_2S, and sulfur dioxide are cooled, and liquid sulfur is recovered and pumped to storage. The other materials are vented to the atmosphere. Besides sulfur produced by the process, high-pressure steam is generated in the waste-heat boiler.

A unit to process 100,000 CF/day of sour natural gas with 15% H_2S has been built. Estimates are available for a commercial plant for treating 8 million CF/day with 15% H_2S. Besides the sweetened and dehydrated gas, the plant would produce 45 tons of sulfur daily.

Another process, by Froning and Grekel, is reported in the patent literature.[21] As with the process by Haines and associates, the sour gas is both dehydrated and sweetened, and elemental sulfur is recovered. However, the regeneration step is carried out with air, which oxidizes the H_2S to free sulfur and water. Two or three beds can be employed. They are used in a cyclic pattern much as described for the other processes.

In the regeneration step, the bed is first preheated by a hot, inert gas. Then air is added to the gas in the bed in an amount sufficient to react with the H_2S, which, at this point, is liberated from the molecular-sieve pores. Sulfur vapor and water vapor are formed and liquid sulfur is recovered. Following regeneration the bed is cooled by passing the sweet-gas stream through it.

The molecular-sieve processes would appear to be most economically attractive for small-to-medium volumes of gas with small H_2S concentrations. Where one can allow the CO_2 to remain in for sales, large CO_2/H_2S ratios make the ability of the processes to selectively remove H_2S even more attractive.

The processes work well at most pressures, but it has been reported[1] that an optimum adsorption pressure for some molecular sieves is about 450 psia. Operating expenses for the molecular-sieve processes are less than those for an iron-sponge unit, and the investment costs are usually less than for an amine unit, particularly if selective sweetening is employed. For larger H_2S concentrations, the prospect of sulfur recovery makes the process even more attractive, but no comparative costs are available.

The main disadvantage of batch operation is negated by the use of smooth, short-cycle operation with two or three towers. Finally, the ability to dehydrate as well as sweeten gas can be a detriment rather than an asset if too much water is present in the gas. Rushton and Hays[61] state that if water content is not below the 20 lb/million SCF required for most efficient H_2S adsorption, it may be necessary to employ upstream dehydration units.

2.1.1.2.3.2. Investment Costs

The investment costs of molecular-sieve gas-sweetening plants for the three different processes summarized below cover a wide range of gas compositions. The wide range of costs is primarily due to the requirement of two, three, or four high-pressure vessels in the cycle, plus the additional control equipment required for the more complex cycles.

	Haines Process[65]	Rushton Unit[61]	EFCO Unit[69]
Gas Rate, 10^6 CF/day	8	1.5	100
Contactor Pressure, psig	700	800	800
H_2S Content,			
in:	15%	25-50 gr/100 SCF	15-20 gr/100 SCF
out:	> 0.25 gr/100 SCF	> 0.25 gr/100 SCF	> 0.25 gr/100 SCF
CO_2 Content,			
in:	--	0.5	6
out:	--	> 0.1	> 0.1
Sulfur Production, tons/day	40	--	--
Regeneration Gas, 1000 CF/hr	6.8	4.2	30
Investment	$600,000	$25,000	$1,000,000

2.1.1.2.3.3. Operating Costs

The operating costs for molecular-sieve gas-sweetening plants vary widely, depending primarily on whether elemental sulfur is to be recovered from the processed gas or the sour gas vented or flared. For the Rushton[61] unit described in the previous section, the operating cost is estimated to be $0.02/1000 CF processed. This is for a plant designed to process 5 million CF/day but presently only handling 1.5 million CF/day. If operated at design conditions, operating cost is estimated to be $0.007/1000 CF.

The Haines[55] Process, with sulfur recovery, is estimated to have about twice the operating cost of the Rushton unit. This is primarily because of the complexity of the operation, although fuel economy is obtained with the sulfur recovery because there is excess heat available, which reduces the energy requirements. This cost is estimated without credit for the recovered sulfur.

The operating cost for the EFCO unit has been estimated[69] at $0.02/1000 CF, with the high cost being due to the removal of carbon dioxide as well as H_2S.

2. 1. 1. 2. 4. Amine Treating – MEA/DEA/TEA

2. 1. 1. 2. 4. 1. Process Description

The commonly used amines are monoethanolamine, diethanolamine, and triethanolamine (MEA, DEA, and TEA) with the more reactive MEA currently being the most widely used. The solutions are generally aqueous, though special solvents have been used for special purposes.

The process (Figure 2.1.1. -18.) is based on the fact that the amines are basic while H_2S is acidic; they react to form salts that decompose readily on heating. The difference in volatility of the amine and H_2S provides for simple separation in the regeneration equipment, the H_2S going overhead as a gas and the lean amine liquid flowing to the bottom of the reactivator. Steam requirements, while not economically prohibitive, still represent a major item of operating cost. Regeneration heat requirements include the heat required to raise the temperature of the rich solution from the absorber temperature to that in the regenerator, the heat of vaporization, and the heat of dissociation of the H_2S amine compounds. Some economy can be effected by heat exchange between the rich and lean solutions. However, only limited benefits can be obtained in this way because of the cost of additional heat exchangers. The higher vapor pressure of MEA solutions as compared to DEA or TEA solutions requires the use of higher regenerator pressures to achieve low H_2S concentrations in the lean solution.

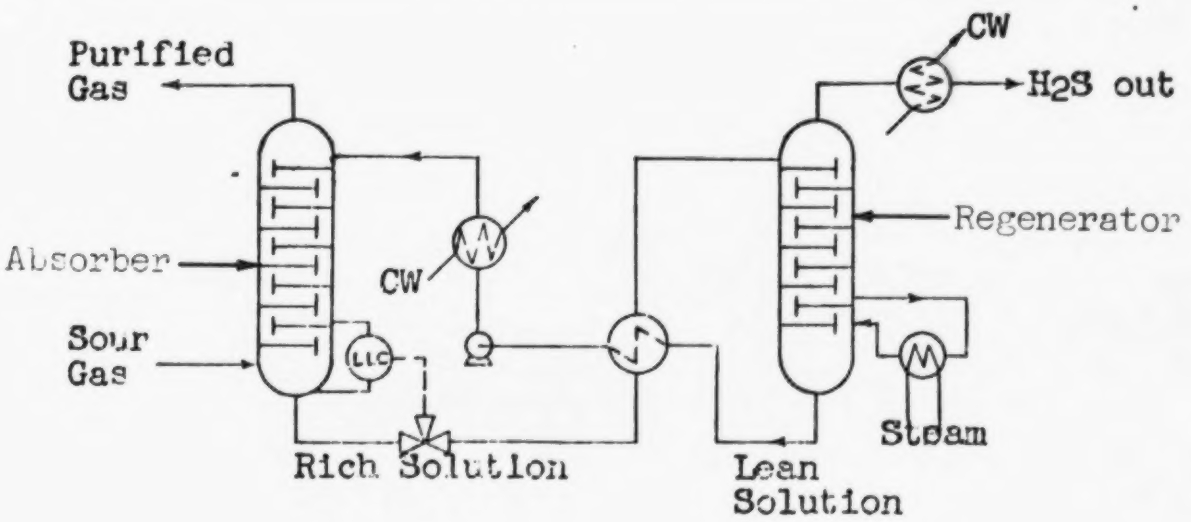


Figure 2. 1. 1. -18. FLOW DIAGRAM OF AMINE TREATING

Alkanolamine, or amine sweetening is the most commonly used of the many processes available.

The principle of the amine process rests on reversible reactions of a weak base with a weak acid to form a water-soluble salt. The reversible nature of these reactions allows regeneration of the amine solution.

These five different types of amine sweetening processes are being or have been used:

- The aqueous MEA process has been most generally used for natural gas sweetening.
- The aqueous DEA process is used primarily for refinery or manufactured-gas sweetening.
- Glycol-amine processes use either DEA or MEA with a glycol to simultaneously sweeten and dehydrate the gas stream.
- Aqueous TEA systems have already been displaced by MEA and DEA systems.
- The methyldiethanolamine (MDEA) system is not commercially competitive with the others

The principal reactions involved in sweetening gas with an aqueous MEA solution are —

$$2RNH_2 + H_2S \rightleftarrows (RNH_3)_2S \qquad (1)$$

$$(RNH_3)_2S + H_2S \rightleftarrows 2RNH_3HS \qquad (2)$$

$$2RNH_2 + CO_2 + H_2O \rightleftarrows (RNH_3)_2CO_3 \qquad (3)$$

$$(RNH_3)_2CO_3 + CO_2 + H_2O \rightleftarrows 2RNH_3HCO_3 \qquad (4)$$

$$2RNH + CO_2 \rightleftarrows RNHCOONH_3R \qquad (5)$$

Similar equations can be written for the other types of amines.

The reactions proceed to the right at low temperatures and to the left at high temperatures. Hence, if the amine contacts the sour gas at a temperature of from 80° to 100°F, the acid gases will be absorbed and react to form the amine salts. If heat is added to this solution and the temperature raised to 240°-250°F, the reaction is reversed, the acid gases displaced, and the amine regenerated for reuse.

Aqueous MEA has found its most general application in the sweetening of wellhead natural gas. It will remove both hydrogen sulfide and carbon dioxide and is generally the first choice when considering an amine sweetening process.

Of the different amines, MEA is the strongest base and hence will react most rapidly with the acid gases. Since it has the lowest molecular weight, it has a greater capacity for acid on a weight basis. This means less solution must be circulated to remove a given concentration of acid gases. Other advantages include its greater stability, its ability to minimize thermal degradation, and the ease with which it can be recovered from contaminated solutions.

Its principal disadvantages are that it will react with carbonyl sulfide and carbon disulfide to form nonregenerative compounds and that its vapor pressure is higher than that of the other amines. A higher vapor pressure can result in greater vaporization losses, but this can usually be overcome with a simple water wash of the sweetened-gas stream.

The MEA process is not considered selective. Although the rate of absorption of CO_2 is less than that of H_2S, the carbon dioxide will be readily absorbed, leaving only small amounts in the sweetened gas.

The MEA process will easily reduce the H_2S content to less than 0.25 gr/100 CF. And if desired, the content can be reduced to less than 0.05 gr/100 CF. This has been accomplished with feed gases containing up to several percent hydrogen sulfide.[42]

The aqueous DEA process finds its most common application in sweetening refinery-gas streams that have been exposed to cracking conditions.

Here carbonyl sulfide and carbon disulfide are usually encountered. Because DEA will not react with these contaminants, it is usually selected for this service.

Hence, the DEA process is usually considered only where the presence of carbonyl sulfide and carbon disulfide eliminate the use of the MEA process. Its only other advantage over the MEA process is that its vapor pressure is lower and hence vapor losses will be less.

Like the MEA process, DEA is nonselective and will remove both H_2S and CO_2. The aqueous DEA process will seldom reduce the hydrogen sulfide content to below 5.0 gr/100 CF.

The aqueous TEA process was the first commercially applied amine sweetening process. Today, it has been largely displaced by either MEA or DEA.

Compared with MEA or DEA, TEA is less reactive with the acid gases, has a smaller capacity per gallon of solution, is less stable, and cannot reduce the H_2S content to some pipeline standards. The principal advantage of the TEA process is the fact that it is selective toward hydrogen sulfide.

In fact, both tertiary amines — TEA and MDEA — are selective toward hydrogen sulfide because they cannot undergo a direct reaction with carbon dioxide (Reaction 5) as they can with hydrogen sulfide. The reactions that do occur involve a water molecule (Reactions 3 and 4) and are much slower than the reactions with hydrogen sulfide.

Aqueous MDEA systems combine the advantages of a tertiary amine — selectivity and lower vapor pressure — with the fact that they are more stable and more reactive than TEA and have a slightly greater capacity. Their principal disadvantage is that they cost 2-4 times as much as the other amines. Today these TEA and MDEA amines find little application as industrial sweetening agents.

The use of mono- or diethanolamine with diethylene glycol has found considerable application in the simultaneous sweetening and dehydration of high-pressure natural gas. The process has two advantages over the aqueous MEA or DEA processes. It will, of course, dehydrate as well as sweeten the gas. In addition, the steam required for the regenerative step is less than for an aqueous solution. The disadvantages include increased amine losses due to vaporization and increased difficulties in purifying the contaminated solution. The use of a glycol-amine system will usually give excellent results in removing acid gases, but generally will not dehydrate as well as a straight glycol plant.

Both MEA and DEA have been used with glycol: MEA because of its greater reactivity and DEA because its boiling point is approximately that of diethylene glycol. The amine choice will govern the degree of sour-gas purification that can be obtained.

Operating experience and consideration of corrosion, circulation rate, vapor pressure, amine costs, and steam requirements have indicated these optimum amine concentrations:

- 15-20% (by weight) for the MEA and DEA systems, although concentrations from 10 to 20% have been used.
- 30% (by weight) solutions of MDEA or TEA have given the best results.
- The glycol-amine system gives good results with a solution of 10-30% (by weight) of amine, 5-10% water, and the balance glycol.

In comparing these processes, some conclusions may be made: In general, the first amine process to consider is the aqueous MEA process. Weighing both its advantages and disadvantages, it seems to be the most economical and most efficient of the five and to offer the best H_2S reduction. In the event carbonyl sulfide and carbon disulfide are present in the sour gas, amine degradation will result.

In this case, it has been found most economical to use the aqueous DEA process. Unfortunately, the DEA process will not purify gas to the same degree as the MEA process.

The glycol-amine process is not as favored as it used to be. Although the glycol will reduce the steam requirements for the sweetening step, dehydration is not as efficient as for a separate glycol plant. The degree of acid-gas purification is, of course, governed by the type of amine used with the glycol. The TEA and MDEA processes would only be considered for a case where selective hydrogen sulfide removal is desired.

A problem affecting the stripping operation of MEA plants has been reported by Fitzgerald and Richardson.[20] For natural gas streams containing unusual proportions of both H_2S and CO_2 (such as ratios of H_2S and CO_2 below 1:70 and greater than 20:1), effective corrosion inhibition may become difficult and problems may occur in obtaining specification residue gas.

Main equipment items in an amine-sweetening plant are the absorber, where the sour gas is sweetened, and the stripper, where the foul amine solution is regenerated. Figure 2. 1. 1.-19 shows a typical amine process.

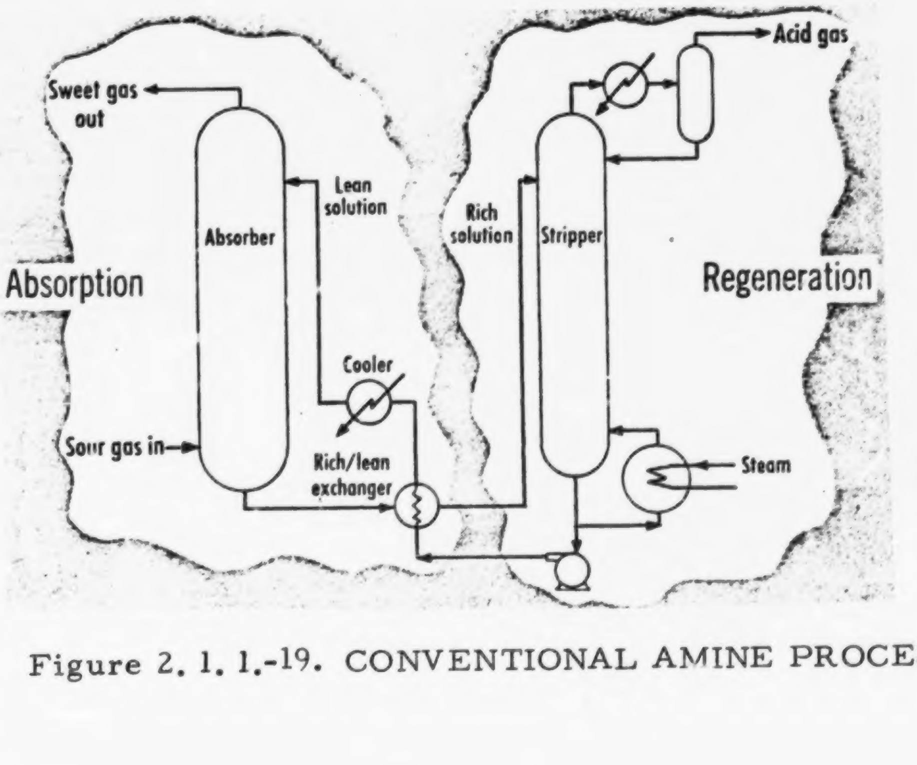

Acid gas
Sweet gas out
Lean solution
Rich solution
Absorber
Stripper
Absorption
Regeneration
Cooler
Sour gas in
Rich/lean exchanger
Steam

Figure 2. 1. 1.-19. CONVENTIONAL AMINE PROCESS [46]

The sour gas containing H_2S and/or CO_2 will usually be passed through a scrubber to remove all free liquids and/or solids, and then will flow upward through the absorber and intimately contact the amine solution. The sweetened gas then leaves the top of the absorber.

The lean amine solution is pumped from the solution surge tank into the top of the absorber and moves countercurrently down the tower, absorbing acid gases from the gas stream. The foul amine solution flows from the bottom of the absorber through a heat exchanger where it is heated by the hot lean solution leaving the regeneration unit.

The solution flows downward through the stripper and vapor generated in the reboiler flows up, stripping the acid gases from the solution. The acid gases and some steam will pass overhead and through a condenser where the steam is condensed and cooled. The gases are separated and sent to a flare or to processing, while the condensate is returned to the top of the stripper as reflux.

The regenerated amine solution flows from the reboiler through the amine heat exchanger where heat is transferred to the cold foul solution. It is further cooled to 80°-100°F before returning to the surge tank.

Several process modifications have been made in an effort to minimize amine losses to the sweetened gas by vaporization. For the aqueous amine processes, a water wash is the obvious answer. This can be done at the top of the absorber using a portion of the acid-gas condensate from the stripper, or water from another source can be used with a drawoff tray to prevent water buildup in the system. Obviously a water wash cannot be used with the glycol-amine system to remove amine vapors if dehydration is desired. One scheme to reduce the amine loss involves a glycol wash. Here, a small stream of solution relatively low in amine content is used to wash the gas stream in the top of the absorber.

Other important equipment, not discussed above but pertinent to an effective amine plant, include a filter and a reclaimer. Filters will remove any solids formed in the solution. Both full-stream filtration and side-stream filtration — where a bypass stream is filtered — have been used. The filter is applied to the lean stream from the stripper.

The reclaimer plays an important role in reducing operating problems and restoring amine usefulness by removing high-boiling and nonvolatile impurities. The reclaimer usually operates on a small side stream (approximately 3%) of lean amine leaving the stripper.

The reclaimer consists of a small batch-distillation vessel. Steam will distill off most of the amine and water, leaving a sludge which is flushed out with water. Different techniques must be used with the different amine processes to prevent thermal degradation of the amine. Reclaiming MEA requires a temperature of approximately 280°-300°F.

For the DEA and TEA systems additional chemicals and equipment are needed — sodium hydroxide, an alcohol to precipitate the salts of the volatile organic acids, and filtration equipment. The glycol-amine reclaimer is even more complicated. The solution is purified by distilling at vacuum and high temperatures (300°-375°F).

A flow modification presented in Kohl and Riesenfeld[42] and shown in Figure 2.1.1.-20 reduces the steam requirement of the aqueous amine processes. The split-stream modification uses two amine solutions fed simultaneously to the absorber: a semilean solution introduced midway down the absorber, and a lean solution introduced at the top.

This process has been used to treat sour gases containing more than 40% acid gas. The semilean solution stream is withdrawn from the midpoint of the reactivation cycle, while the thoroughly regenerated lean solution is removed from the bottom of the stripper. The obvious drawback of this modification is that it appreciably increases the initial cost of the treating plant.

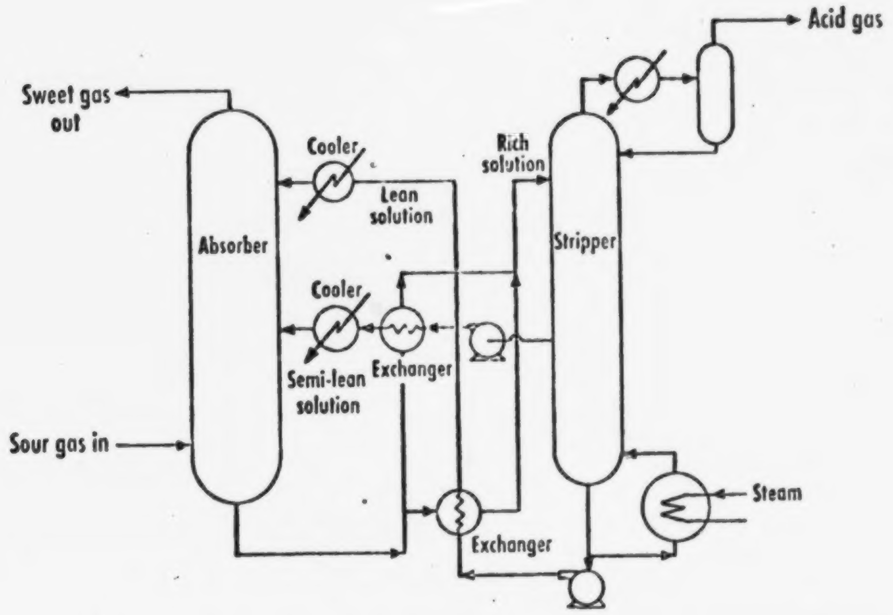

Acid gas
Sweet gas out
Cooler
Lean solution
Rich solution
Absorber
Stripper
Cooler
Exchanger
Semi-lean solution
Sour gas in
Steam
Exchanger

Figure 2. 1. 1.-20. SPLIT-STREAM AMINE PROCESS[49]

2. 1. 1. 2. 4. 2. Investment Costs

Estimates[60] of treating plant costs have been updated to 1968 prices.[54] The particular example given is for monoethanolamine. However, costs for systems using other amines would be comparable.

Inlet Gas Flow, 10^6 SCF/day	90
Contactor Operating Pressure, psig	1000
CO_2 Concentration,	
in, %	28.0
out, ppm	10-25
H_2S Concentration,	
in, gr/100 SCF	2
out, gr/100 SCF	<0.25
Total Investment, $	4,150,000

For this process, the investment costs depend preimarily on the total gas throughput, the absorber operating pressure, the percentage of acid gas in the inlet stream, and the required minimum outlet concentrations to meet contract specifications. Also, for gases containing H_2S, the carbon dioxide content in the exit gas must be much lower than normally required if an acceptably low H_2S concentration is to be reached.

2.1.1.2.4.3. Operating Costs

Amine plant operating costs are principally a function of acid-gas concentration in the feed gas, the degree of removal, plant size, and operating pressure. The principal items of operating cost are regeneration steam, amine pump power, and makeup chemicals. Typical operating costs for a natural gas sweetening plant treating a low-CO_2-content gas are —

	¢/1000 CF
Steam (at $0.55/1000 lb)	0.18
Energy (at $0.01/kWhr)	0.07
Chemicals (Including Process Water)	0.10

2. 1. 1. 2. 5. Glycol-Amine

The glycol-amine process was developed and used where simultaneous acid-gas removal and dehydration of natural gas was desired. This process is one of the most widely used techniques for treatment of natural gas prior to gas pipeline transmission.

2. 1. 1. 2. 5. 1. Process Description

The general process steps for the glycol-amine system were described in 2. 1. 1. 1. 2. 1. and will only be summarized here. This process uses a mixture of monoethanolamine (MEA), diethylene glycol (DEG), and water. Typical compositions are from 10 to 30 weight percent MEA, 45 to 85 weight percent DEG, and 5 to 25 weight percent water.

The glycol-amine solution dehydrates gas to a slightly lower moisture dew point than a DEG-water solution having the same water concentration. The MEA appears to combine with the same quantity of acid gas and to achieve the same reduction in concentrations as the equivalent MEA-water solution. Thus, the solution circulation rate for the glycol-amine process is 2 to 3 times that required by the amine-water system for the same quantity of acid gas removed. However, the flexibility inherent in the glycol-amine process to simultaneously achieve a wide range of acid gas and water removal in the treated gas frequently makes this process the economic choice.

Very high purity gas can be obtained. Typical field data from a plant operating with the contactor at 550 psig, with inlet gas containing 126 gr H_2S/100 SCF and 6.24 mole percent CO_2, gave an outlet gas with 0.02 gr H_2S/100 SCF and less than 0.01 mole percent CO_2

2.1.1.2.5.2. Investment Costs

Estimates[35] for a natural gas processing plant using MEA and DEG have been updated to 1968 prices.[54] Since this plant was to be added to an existing installation, utilities were already in existence on the site. The operating conditions of the plant were –

Inlet Gas Flow, 10^6 SCF/day	7.5
Contactor Operating Pressure, psig	375
Inlet CO_2 Concentration, %	5.00
Outlet CO_2 Concentration, %	0.15
Water Content,	
in	Saturated at separator conditions
out, lb/10^6 SCF	10
Total Investment, $	395,000

2.1.1.2.5.3. Operating Costs

The principal operating costs for glycol-amine plants, other than operating labor costs, are for the steam required for regeneration, pumping energy for the liquid absorbent, and the chemicals to replace losses. For a typical small plant these costs have been estimated as –

	¢/1000 CF Treated
Steam (at $0.55/1000 lb)	0.16
Energy (at $0.01/kWhr)	0.03
Chemicals (including process water)	0.01

To these costs must be added maintenance costs, usually estimated at about 4%/yr; labor costs, which are highly dependent upon the degree of automation; and overload costs.

2.1.1.2.6. Giammarco-Vetrocoke

2.1.1.2.6.1. Process Description

There are two distinct processes involved in the Giammarco-Vetrocoke process, one for the removal of CO_2 from gases free of H_2S and another for selective H_2S removal. These processes are based on the absorption of the acid gases in an alkaline solution containing inorganic or organic additives. Typical inorganic additives used in the CO_2 absorption process are arsenic oxide and selenious and telluious acid. Sodium and potassium carbonate solutions containing trivalent and pentavalent arsenic are used in the H_2S-removal process.

The processes have been applied in Europe in several commercial installations for the purification of coke-oven and synthesis gases.

A typical flow diagram of a carbon dioxide removal process is shown in Figure 2.1.1.-21. In this case, which is typical for the treatment of a gas stream containing only carbon dioxide, the solution is regenerated with steam.

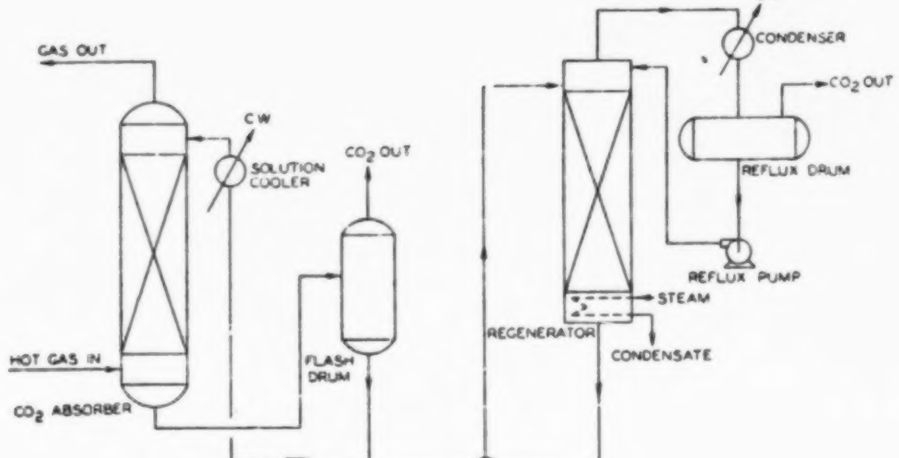

GAS OUT
CW
SOLUTION COOLER
CO_2 OUT
CW
CONDENSER
CO_2 OUT
REFLUX DRUM
REFLUX PUMP
STEAM
REGENERATOR
CONDENSATE
HOT GAS IN
FLASH DRUM
CO_2 ABSORBER
SOLUTION PUMP
HEAT EXCHANGER

Figure 2.1.1.-21. GIAMMARCO-VETROCOKE CO_2 REMOVAL PROCESS WITH STEAM REGENERATION [60]

The CO_2-containing gas enters at the bottom of the absorber, which may be any contacting device ensuring intimate countercurrent contact between the gas and the liquid, and the purified gas leaves the absorber at the top. The lean solution enters the absorber at the top and flows from the bottom of the absorber to the flash drum where the pressure is reduced and an appreciable portion of the carbon dioxide is released. The partially stripped solution flows from the bottom of the flash drum through a heat exchanger, where it is heated by exchange with the regenerated solution, to the top of the regenerator, where the remaining carbon dioxide is stripped from the solution. Heat is supplied to the bottom of the regenerator by means of indirect steam. The lean solution leaves the bottom of the stripper and, after heat exchange with the incoming rich solution, is returned to the top of the absorber.

The process scheme, including air regeneration,is shown in Figure 2.1.1.-22which is a typical flow diagram for the treatment of a gas stream containing both hydrogen sulfide and carbon dioxide. The gas is first passed through an absorber where the hydrogen sulfide is removed, and then enters the absorber of the CO_2-removal section. The H_2S-removal process is discussed in detail in a later portion of this section. The flow scheme of the CO_2-removal section is very similar to that described above. The lean solution enters at the top of the absorber, and carbon dioxide is removed in a countercurrent operation. The rich solution leaving the bottom of the absorber is heated before it enters the flash drum. From the flash drum, the partially stripped solution flows to the top of the regenerator.

Final stripping is obtained by contact with a stream of preheated and water-saturated air which enters at the bottom of the regenerator. The mixture of carbon dioxide and air leaving the top of the regenerator is washed with a stream of cold water in the dehumidifier to recover heat and condense water contained in the gas stream. The cooled mixture of air and CO_2 leaving the top of the dehumidifier is vented to the atmosphere. The water stream leaving the bottom of the dehumidifier is used to presaturate and preheat the incoming air in the presaturator. This arrangement results in a maximum of heat recovery and a minimum of water loss.

The H_2S-removal process is based on the absorption of hydrogen sulfide in solutions containing a mixture of alkaline arsenites and arsenates. Considerable flexibility, such as selective H_2S absorption or simultaneous absorption of both H_2S and CO_2, can be obtained by proper selection of the proportion of arsenite to arsenate and the pH of the solution.

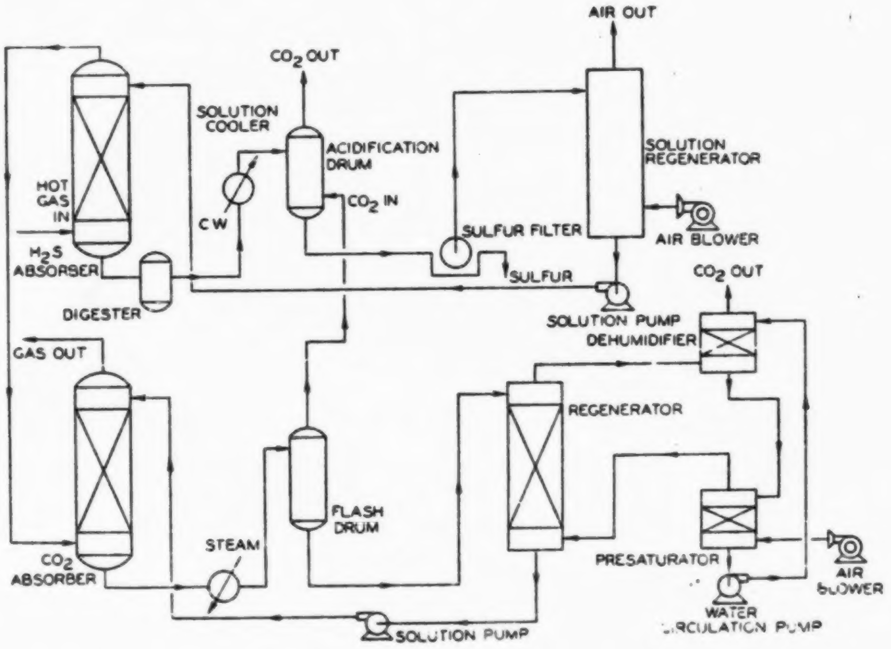

AIR OUT
CO_2 OUT
SOLUTION COOLER
ACIDIFICATION DRUM
SOLUTION REGENERATOR
HOT GAS IN
CW
CO_2 IN
SULFUR FILTER
AIR BLOWER
H_2S ABSORBER
SULFUR
CO_2 OUT
DIGESTER
SOLUTION PUMP
DEHUMIDIFIER
GAS OUT
REGENERATOR
FLASH DRUM
STEAM
CO_2 ABSORBER
PRESATURATOR
AIR BLOWER
WATER CIRCULATION PUMP
SOLUTION PUMP

Figure 2.1.1.-22. GIAMMARCO-VETROCOKE H_2S- AND CO_2-REMOVAL PROCESS, INCLUDING AIR REGENERATION[60]

The flow scheme of the process is shown diagrammatically in Figure 2.1.1.-21, and 2.1.1.-22. The scheme shown in Figure 2.1.1.-22 includes acidification with carbon dioxide under pressure, while this step is omitted in the version presented in Figure 2.1.1.-23. The H_2S-containing gas (Figure 2.2.2.-22) is fed to the bottom of the absorber, which may be any efficient liquid-gas contacting device, where it countercurrently contacts the lean solution entering at the top of the vessel. The rich solution is withdrawn at the bottom of the absorber and flows to a surge vessel called the digester, where the relatively slow conversion reaction of thioarsenite to monothioarsenate is completed. The solution then flows to the acidification vessel where it is treated with a stream of carbon dioxide. In the scheme shown in Figure 2.1.1.-21, the carbon dioxide is taken from the flash

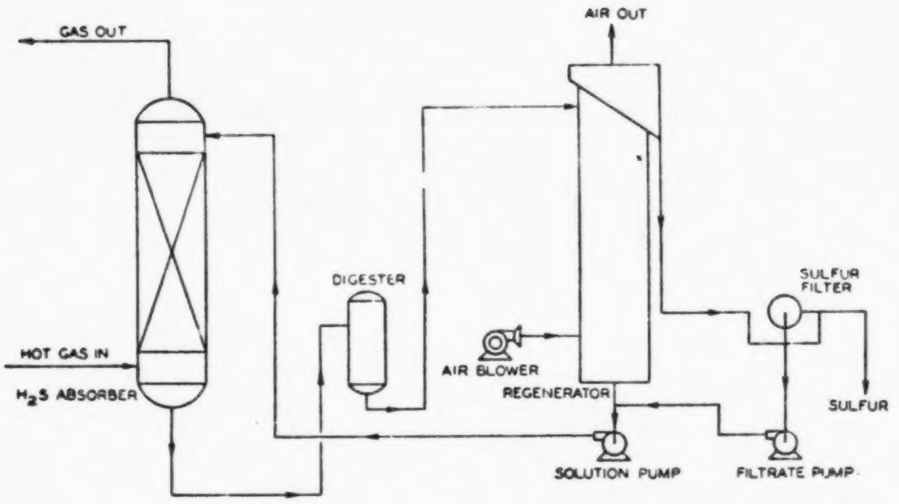

GAS OUT
AIR OUT
DIGESTER
SULFUR FILTER
HOT GAS IN
H_2S ABSORBER
AIR BLOWER
REGENERATOR
SULFUR
SOLUTION PUMP
FILTRATE PUMP

Figure 2.1.1.-23. GIAMMARCO-VETROCOKE H_2S-REMOVAL PROCESS WITH AIR REGENERATION [60]

drum of the CO_2-removal section which follows the H_2S-removal plant. The slurry containing elemental sulfur is withdrawn from the acidifier, the sulfur is removed by filtration, and the solution is fed to the top of the oxidizer where trivalent arsenic is reoxidized to pentavalent arsenic by contact with air.

2. 1. 1. 2. 6. 2. Investment Costs

Three estimates have been reported for the investment costs of the Giammarco-Vetrocoke Process for removing carbon dioxide and/or hydrogen sulfide from natural gas.[35, 60] These estimates include regeneration by both air stripping and steam stripping. The costs have been adjusted to 1968 values by use of the "Nelson Cost Indexes" refinery index.[54]

Commercial experience has shown that this process offers definite investment cost advantages over other systems when large percentages of carbon dioxide and low-hydrogen-sulfide-content gases are to be treated. A final cleanup with MEA or by adsorption is required to reduce the carbon dioxide content low enough for liquefaction of the natural gas. Investment costs for three plants based on this process are given below. The Case I costs also include a TEG dehydration unit which is not included in Cases II and III.

	Case I	Case II	Case III
Regeneration by	Air stripping	Air stripping	Steam stripping
Inlet, 10^6 SCF/day	180	90	90
Contactor, psig	1050	1000	1000
% CO_2,			
in:	28	28	28
out:	2	0.5	0.1
H_2S gr/100 SCF,			
in:	2.5	2	2
out:	"Sweet" to lead acetate paper	<0.25	<0.25
Total Investment, $	5,100,000	2,800,000	2,250,000

Investment costs for this process are primarily dependent on the total gas throughput rate, the absorber operating pressure, the acid-gas concentration in the inlet stream, and the required outlet concentration (the lower the inlet/outlet ratio, the lower the cost of the plant).

2.1.1.2.6.3. Operating Costs

Operating cost estimates have been reported for three plants based on this process. The operating conditions for each plant and the assumptions used in calculating the cost are included for each case. The principle operating costs, regeneration energy (either steam or air in this process), and absorbate pumping energy are given.

	Case I	Case II	Case III
Regeneration by	Steam stripping	Air stripping	Steam stripping
Inlet, 10^6 SCF/day	7.5	90	90
Contactor, psig	375	1000	1000
% CO_2,			
in:	5.0	28.0	28.0
out:	0.15	0.1	0.5
H_2S gr/100 SCF,			
in:	--	2	2
out:	--	>0.25	>0.25
Steam (at $0.55/1000 lb), ¢/1000 CF	0.86	0.34	1.5
Energy (at $0.01/kWhr), ¢/1000 CF	0.25	0.44	0.3
Chemicals, ¢/1000 CF	0.06	0.04	0.1

2.1.1.2.7. Hot Potassium Carbonate Process

2.1.1.2.7.1. Process Description

The hot potassium carbonate system is similar in many respects to an amine unit. The major items of equipment such as the contactor, regenerator, circulating pumps, and reboiler surface are essentially the same. Since absorption and regeneration steps both occur at the same temperature in the hot carbonate process, no liquid heat exchangers are required.

The advantages of this process are best realized at pressures above 250 psi, with gas compositions from 50-5% acid gas on the inlet, and with extraction of the acid gases to a purity of 0.5%. A greater purity will require special considerations to reduce the H_2S content to less than 0.25 gr/100 SCF: the common pipeline specification for natural gas.

In a natural gasoline plant the logical location for the process is on the compressor discharge before cooling and dehydration. Additional heat for regeneration can be derived from various sources of waste, for example, the still reflux systems, hot lean oil, and jacket water heat.

Figure 2.1.1.-24 shows the components of a hot carbonate plant. Hot or cold acid-bearing gases enter the base of the contactor where they countercurrently contact the carbonate solution. Treated gas, leaving the top of the contactor, is saturated with water and is cooled prior to delivery to a pipeline for dehydration, and water recovery. Carbonate, the continuous phase, is pumped to the top of the contactor and trapped from the bottom by a liquid-level control. It passes directly to the regenerator where the heat of decomposition is added. Carbon dioxide leaves the top of the stripper through a back-pressure regulator. Water carried out by the carbon dioxide may be recovered in a condenser or vented, depending on the value of water. This particular scheme is best suited for removal of acid gases from streams containing 5% or higher of acid components. The practical limit of residue purity is about 1%.

Figure 2.1.1.-25 shows a split-stream carbonate process that is applicable to natural gas having a very high acid-gas content. A partially stripped stream of hot carbonate is fed to the lower section of the contactor where the acid-gas partial pressure is high. The remaining portion of the carbonate is thoroughly regenerated in the reboiler and returned to the top part of the contactor where the final extraction takes place. This method is desirable for a minimum fuel requirement where economics justify the extra equipment. The same practical limits as in the basic carbonate unit apply.

The flow diagram in Figure 2.1.1.-26 shows a combination water absorber and a hot carbonate process. This scheme is only applicable to very high acid-gas streams with carbon dioxide content near 50%. Because of the high hydrocarbon absorption, this process is usually unattractive, but, where a large quantity of low Btu fuel (300 Btu) can be used, it may be acceptable.

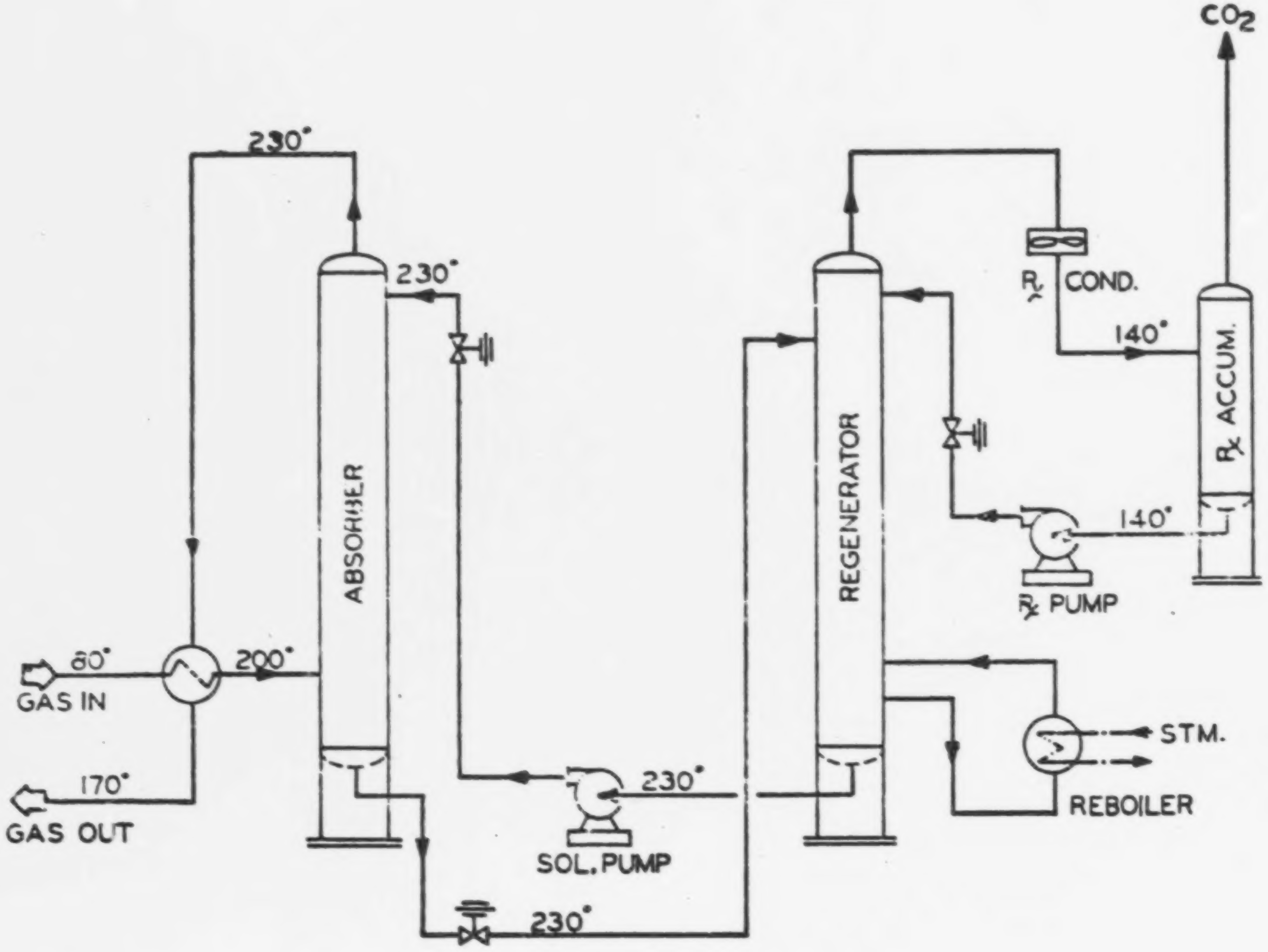

CO2
230°
230°
ABSORBER
REGENERATOR
R_x COND.
140°
R_x ACCUM.
140°
R_x PUMP
80°
GAS IN
200°
170°
GAS OUT
230°
SOL. PUMP
230°
STM.
REBOILER

Figure 2.1.1-24. HOT POTASSIUM CARBONATE ACID-GAS REMOVAL PLANT[7]

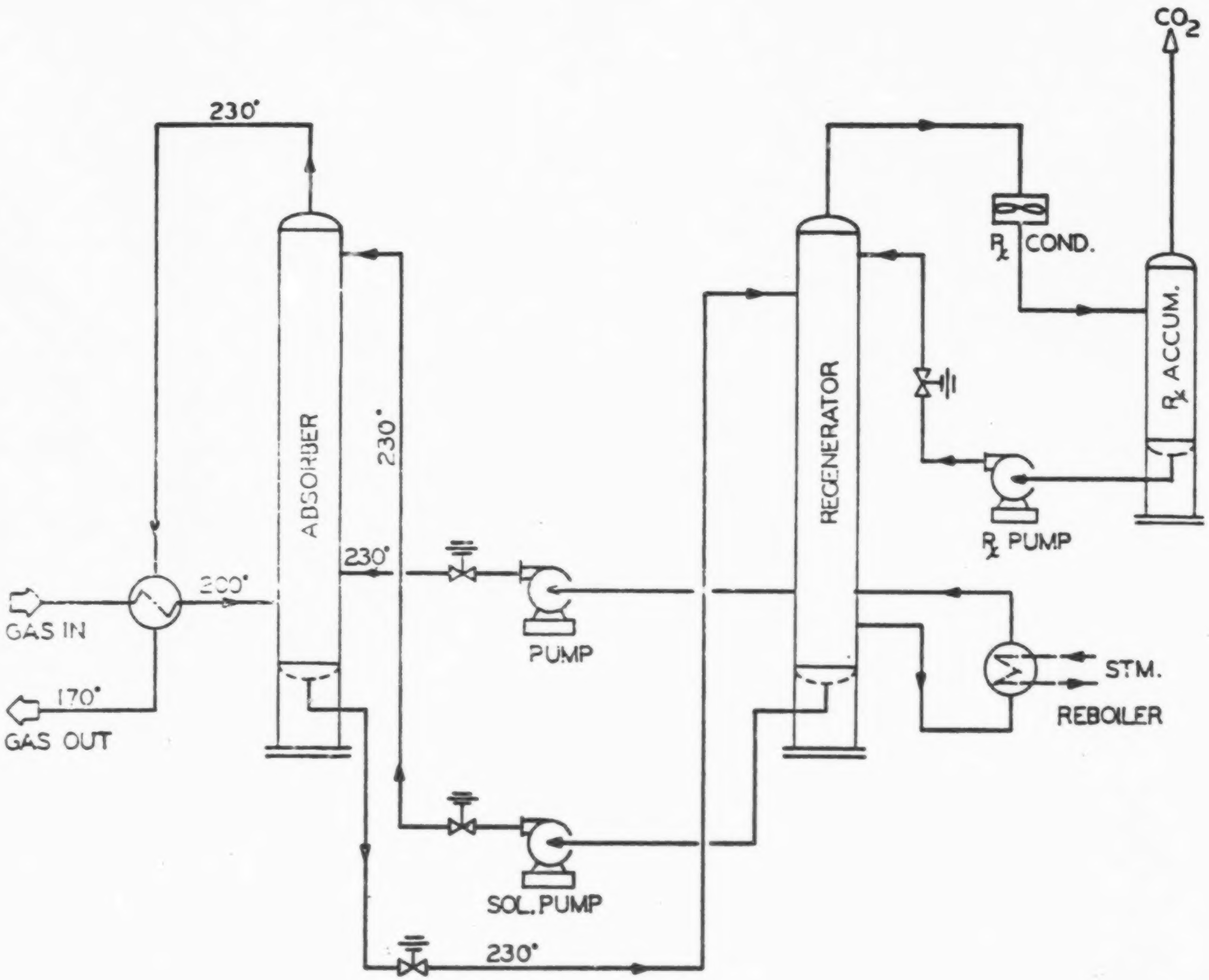

CO$_2$
230°
R_x COND.
R_x ACCUM.
ABSORBER
230°
REGENERATOR
R_x PUMP
230°
GAS IN
200°
170°
GAS OUT
PUMP
STM.
REBOILER
SOL. PUMP
230°

Figure 2.1.1.-25. SPLIT-STREAM HOT CARBONATE SYSTEM FOR ACID-GAS REMOVAL[7]

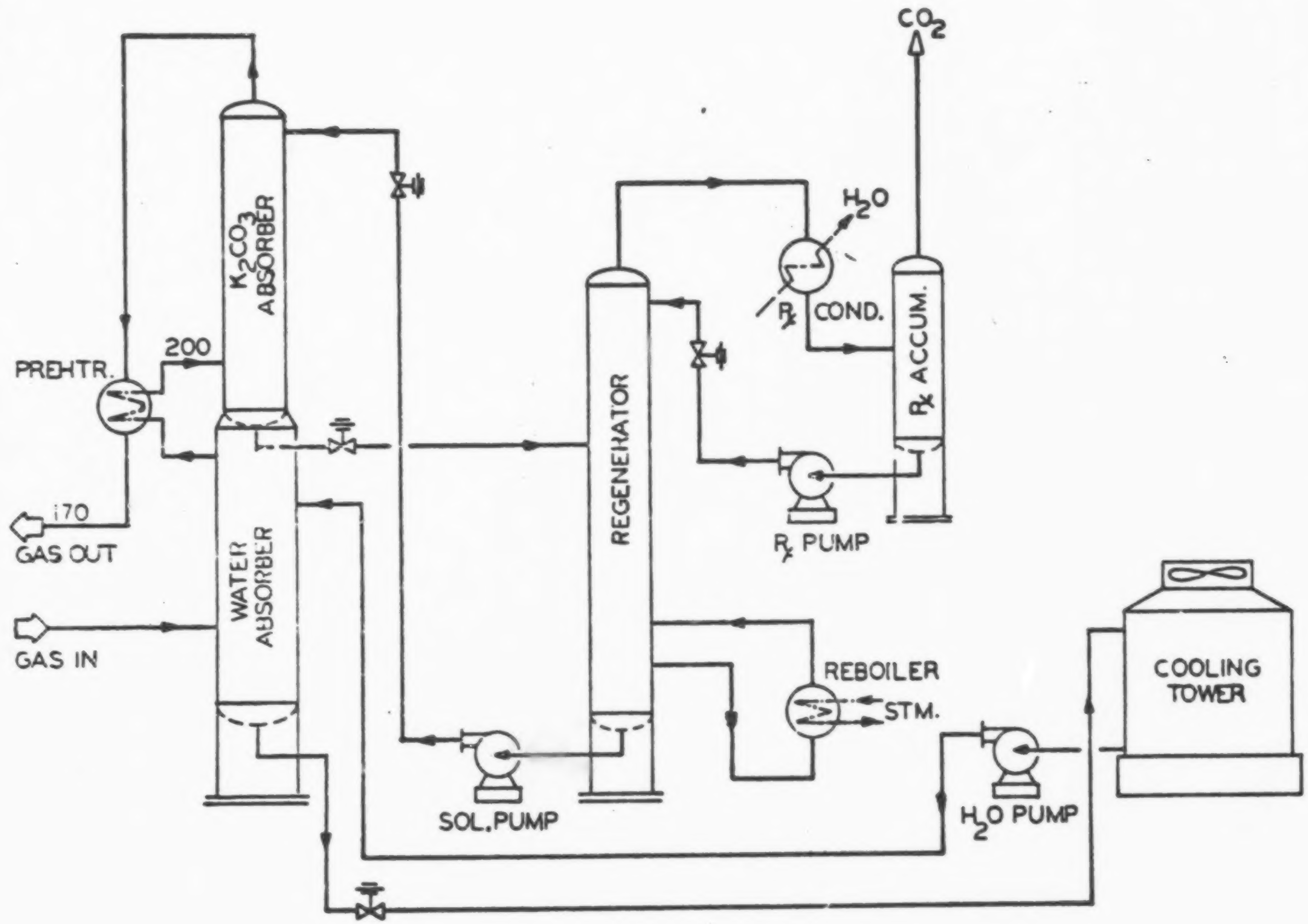

CO2
H2O
R_x COND.
R_x ACCUM.
R_x PUMP
K_2CO_3 ABSORBER
PREHTR.
200
170
GAS OUT
GAS IN
WATER ABSORBER
REGENERATOR
REBOILER
STM.
SOL. PUMP
H_2O PUMP
COOLING TOWER

Figure 2.1.1.-26. WATER CARBONATE PROCESS FOR ACID-GAS REMOVAL[7]

Figure 2.1.1.-26a shows the same process as Figure 2.1.1.-24 with the addition of a lean carbonate cooler for a portion of the stream. Cooling the carbonate decreases the equilibrium carbon dioxide pressure in the carbonate for use at the top of the contactor. This scheme is used to reduce the off-gas to a low acid gas specification. If the circulation of the cooled solution is fast enough, this scheme can eliminate a gas cooler. The practical limit for this process is about 0.1% acids in the residue. This scheme may be combined with the split-flow scheme of Figure 2.1.1.-25 to advantage. A modification of this system in which a small stream of potassium hydroxide is fed to the top of the contactor to react with the carbon dioxide or hydrogen sulfide will result in high residue purity. The reacted bicarbonate or bisulfide is fed to the system as makeup.

2.1.1.2.7.2. Investment Costs

The economies of a hot potassium carbonate acid-gas removal unit depend a great deal on the type of gas to be treated, the pressure of the gas, and the concentration of acid gas. A stringent specification will cost more to meet than a 5% residue specification.

Figure 2.1.1.-27 shows the spread between what we have termed a basic carbonate unit, where gas is delivered hot to the system and is returned saturated with water vapor, and a total grass roots treating unit, where gas is received at ambient temperatures and delivered to a pipeline, sweet and dry.

It is apparent that, as the plant capacity is increased, the unit investment for a removal system is decreased markedly. This is primarily due to the inherent savings in a large-capacity installation coupled with an assumption that the larger capacity was partly due to an increased concentration as well as higher gas rates.

2.1.1.2.7.3. Operating Costs

The most marked economy in the use of a carbonate system is operating costs. For example, operating costs involved in the purification to 1% of 25 million SCF/day of natural gas containing 50, 25, and 2% acid gas with no capital depreciation charge are estimated at \$0.058, \$0.029, and \$0.013/1000 SCF. This charge includes the estimated costs of fuel and utilities,

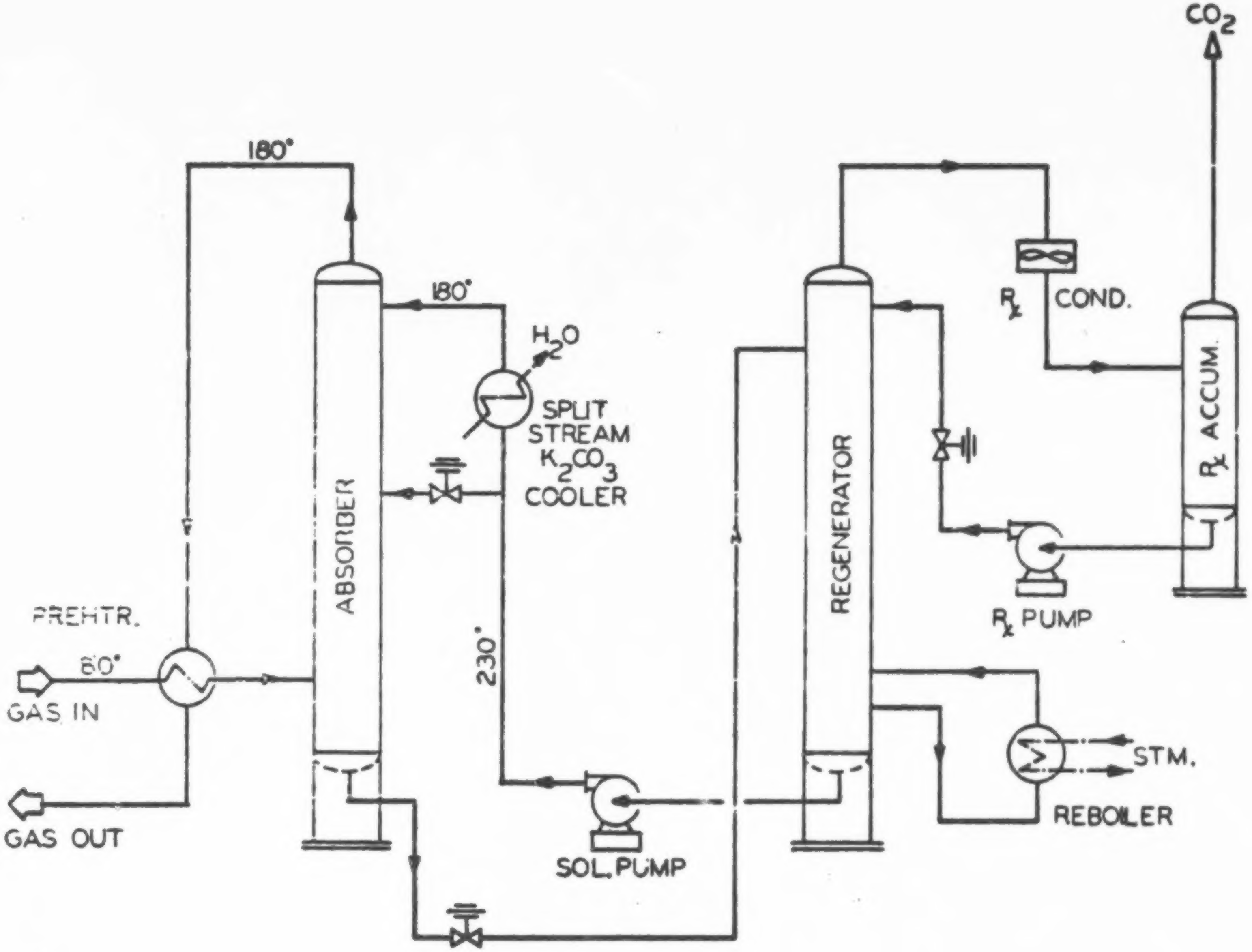

CO2
180°
180°
H2O
SPLIT STREAM K2CO3 COOLER
Rx COND.
Rx ACCUM.
ABSORBER
REGENERATOR
Rx PUMP
PREHTR.
60°
GAS IN
GAS OUT
230°
STM.
REBOILER
SOL. PUMP

Figure 2.1.1-26a. HOT CARBONATE PROCESS WITH COOLED SPLIT-STREAM FOR ACID-GAS REMOVAL 7

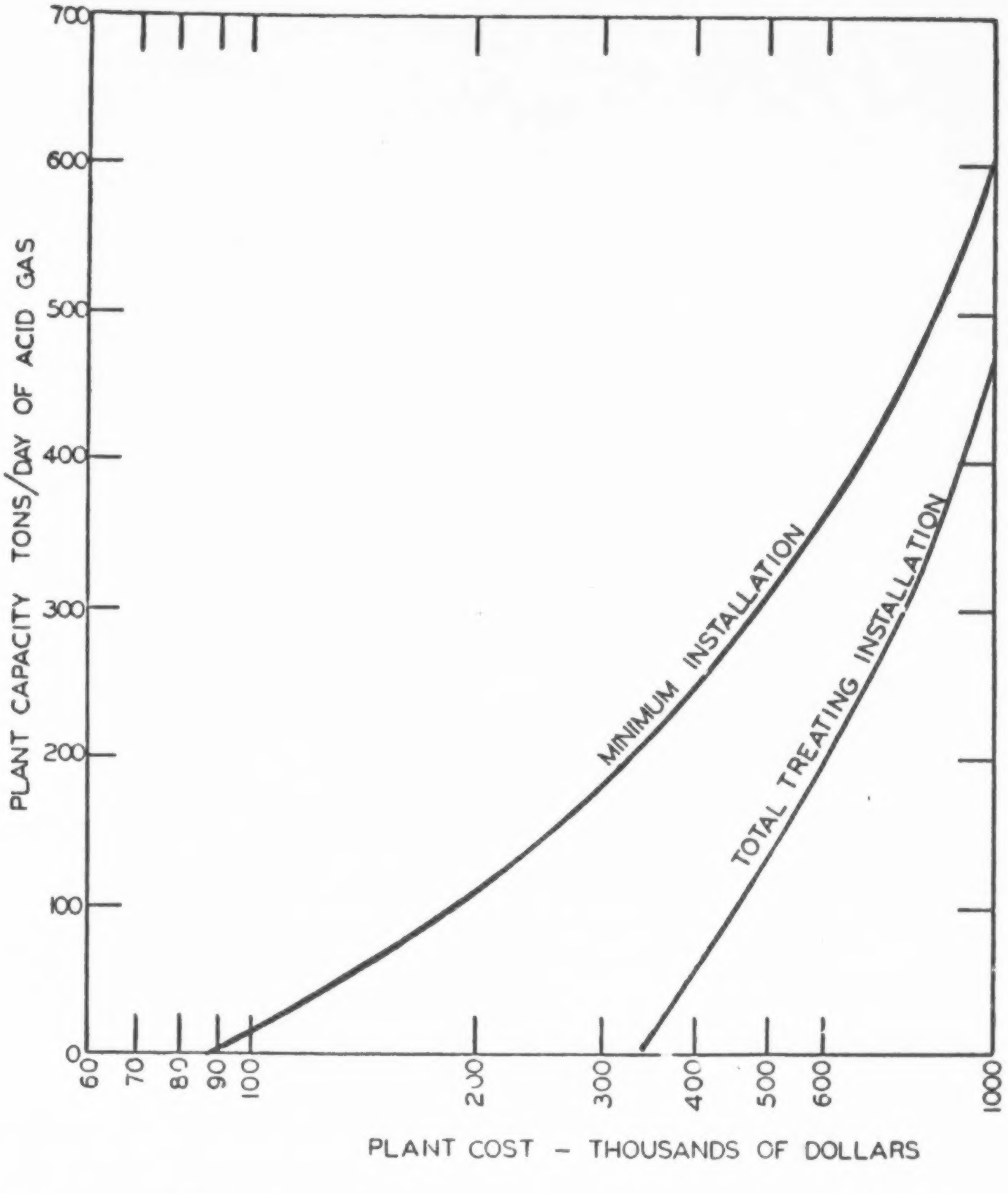


Figure 2.1.1.-27. ANALYSIS OF PLANT COSTS BY ACID-GAS CAPACITY[7]

water system, steam plant or heaters, chemicals, and operating labor. Gas is priced at 12¢/100 SCF, electric power at 1¢/kWhr, and labor at $120/day. These charges are for a grass roots installation where operating labor or supervisory labor are not already available

2.1.1.2.8. Sulfur Recovery — Townsend

2.1.1.2.8.1. Process Description

The principal features of this process are —

a. Simultaneous gas sweetening and sulfur recovery
b. Insensitivity to carbon dioxide in the process gas stream
c. Effective dehydration of the residue gas
d. Over 95% of the sulfur, in the gas as H_2S, recovered as sulfur

The limitations are —

a. Does not strip carbon dioxide
b. Efficiency decreases with increasing gas stream water content
c. Does not strip nitrogen and helium

A typical Townsend Process is shown in Figure 2.1.-6. Sour natural gas enters the lower portion of the reactor at line pressure and atmospheric temperature and rises through the reactor column, where it contacts a descending stream of 98% glycol containing the proper amount of sulfur dioxide, which is pumped into the top of the reactor.

When the gas and liquid phases contact, the reaction is virtually instantaneous; sulfur and water are formed. The sulfur is insoluble in the glycol, but the water is absorbed. Glycol is used in sufficient quantity that water concentration increases to 5% by weight or less after absorbing all water produced by the reaction plus the water vapor content of the inlet gas.

Sweetened gas rises from the reactor and carries with it some excess sulfur dioxide. This mixture enters the base of the high-pressure absorber, where it contacts a descending stream of concentrated, sulfur-dioxide-free glycol that is introduced at the top of the column. This solution absorbs the excess sulfur dioxide and the remaining water vapor in the gas and passes downward into the reactor as shown in Figure 2.1.1.-28. Sweet, dry gas leaves the top of the absorber.

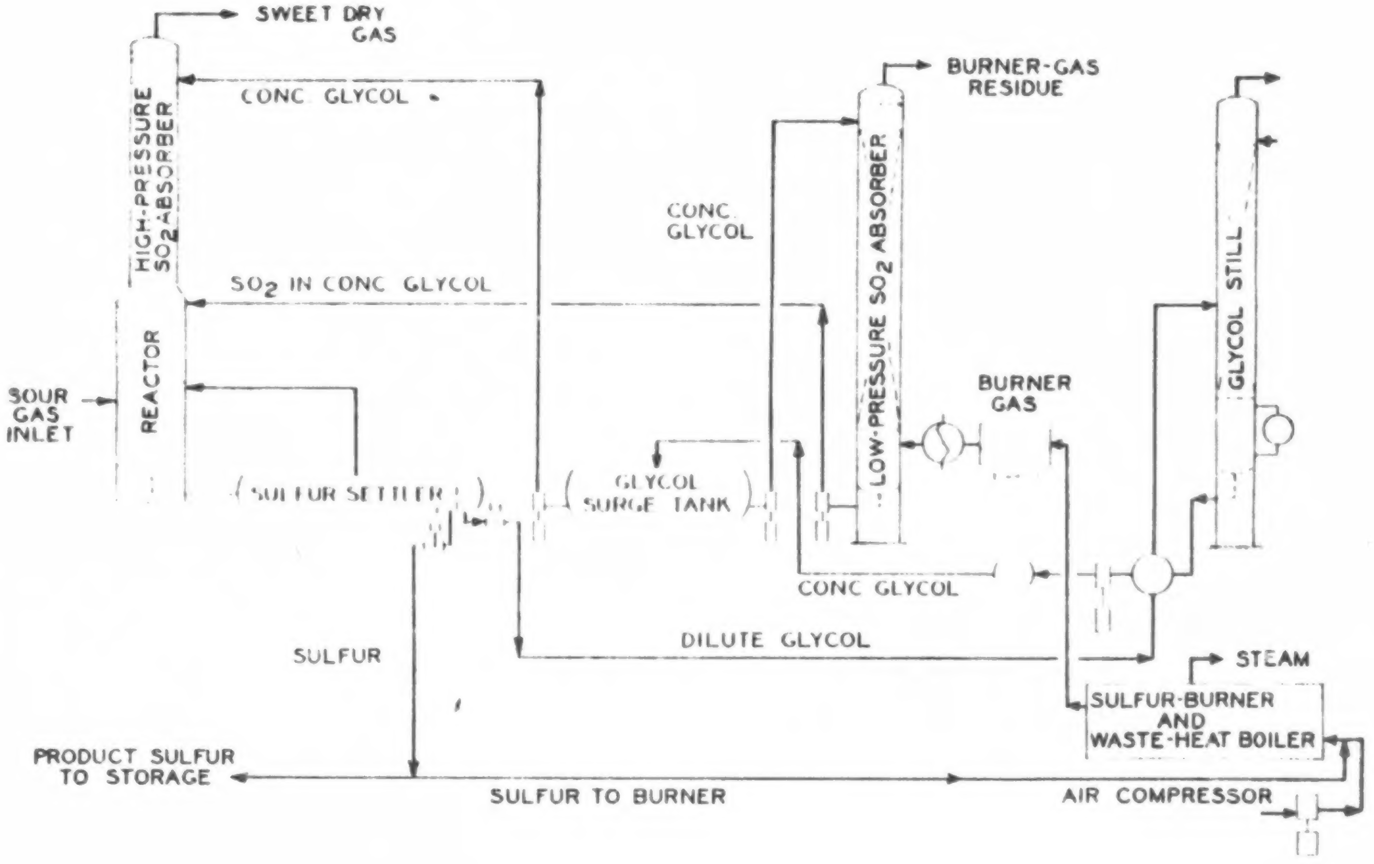

SWEET DRY GAS
HIGH-PRESSURE SO2 ABSORBER
CONC GLYCOL
BURNER-GAS RESIDUE
CONC GLYCOL
LOW-PRESSURE SO2 ABSORBER
GLYCOL STILL
SO2 IN CONC GLYCOL
SOUR GAS INLET
REACTOR
BURNER GAS
SULFUR SETTLER
GLYCOL SURGE TANK
CONC GLYCOL
DILUTE GLYCOL
SULFUR
STEAM
SULFUR-BURNER AND WASTE-HEAT BOILER
PRODUCT SULFUR TO STORAGE
SULFUR TO BURNER
AIR COMPRESSOR

Figure 2.1.1.-28. PIPELINE GAS SERVICE FLOW DIAGRAM OF TOWNSEND SULFUR-RECOVERY PROCESS

Another variation of this process is shown in Figure 2.1.1.-29 as an example.

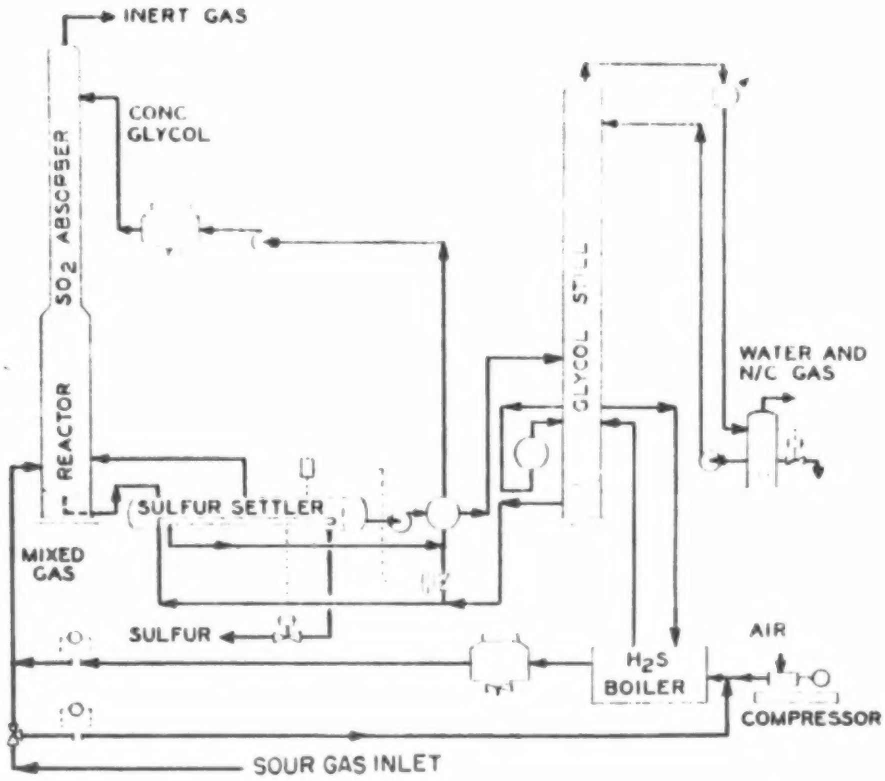

INERT GAS
CONC GLYCOL
SO2 ABSORBER
REACTOR
GLYCOL STILL
WATER AND N/C GAS
SULFUR SETTLER
MIXED GAS
SULFUR
AIR
H2S BOILER
COMPRESSOR
SOUR GAS INLET

Figure 2.1.1.-29. ACID-GAS SERVICE FLOW DIAGRAM OF TOWNSEND SULFUR-RECOVERY PROCESS

The Townsend Process may be employed to recover sulfur from acid gases produced by an amine-type sweetening plant. The sour gas stream is split so that one portion, comprising one-third of the total stream, is mixed with air and burned in a boiler to produce the sulfur dioxide requirement for the process.

Heat from the boiler may be used to generate process steam or to reconcentrate the glycol solution employed in the process. Flue gas from the boiler is cooled to 120°-130°F prior to mixing with the remaining two-thirds of the acid-gas stream. The commingled gas is then charged to the reactor. It is important to note that it contains the reactants in stoichiometric proportion and that the excess sulfur dioxide required for complete conversion of the hydrogen sulfide is retained in and recycled between the reactor and the absorber — a feature that contributes substantially to the improvement of sulfur-recovery efficiency.

The reaction between hydrogen sulfide and sulfur dioxide proceeds in the reactor, as described before. Sulfur is melted and withdrawn from the system and glycol is decanted, reconcentrated, and recirculated in precisely the same manner shown in Figure 2.1.1.-29. Freshly reconcentrated glycol exchanges heat with the contents of the settling drum and with the glycol feed to the still. Then it is cooled to near-atmospheric temperature and pumped to the top of the absorber, where it takes up excess sulfur dioxide and water vapor as it descends into the reactor.

2.1.1.2.8.2. Investment Costs

The Townsend Process accomplishes in one plant what previously required two or three separate units: gas sweetening, residual gas dehydration, and sulfur recovery. Due to the absorbent used, dehydration is inherent in the process. Sulfur recovery efficiency is considered better than that obtained with the 2-stage Claus-Chance Process.

Investment costs for the Townsend Process are estimated to be 5-10% lower than those for the Claus-Chance Process, when considered as a treating plant for acid gases. When considered for simultaneous dehydration, acid-gas removal (H_2S only), and sulfur recovery, the Townsend Process appears to have a distinct economic advantage.

2.1.1.2.8.3. Operating Costs

Power consumption for the Townsend Process is less than for separate sweetening, dehydration, and sulfur recovery plants. Labor requirements are about the same as for amine-water sweetening plants. If the gas to be processed contains only a few grains of H_2S/100 SCF, sulfur is required as an external source to provide the necessary SO_2 makeup. However, as the H_2S concentration in the inlet gas increases, the process becomes a net sulfur producer.

Economic studies for relatively large plants comparing the amine-water and Townsend processes indicate the operating costs should be almost identical when CO_2 concentration is 1%. For higher CO_2 concentrations, the Townsend Process is favored (if it is not necessary to remove the CO_2). Depending on the market price of sulfur, the break-even point for combined operating and capital charges will be for H_2S concentrations between 2 and 3%.

However, there is an operating difficulty that has precluded use of the Townsend Process in the U.S. to date. The presence of organic sulfides in natural gas apparently causes the formation of noxious thio-compounds when treated by the Townsend Process; this difficulty has not yet been eliminated.

2. 1. 1. 2. 9. Sulfur Recovery — Claus-Chance

2. 1. 1. 2. 9. 1. Process Description

The conversion of H_2S to elemental sulfur provides only a small portion of the industrial sulfur supply, but this source has the advantage over most other secondary sulfur sources of producing a very high purity sulfur, actually 99.9+%, and negligible ash from combustion. This is better than the elemental sulfur produced by the Frasch Process operating near the Gulf Coast. Furthermore, sour-gas fields are found in many different localities, giving sulfur produced from H_2S a distinct marketing advantage.

The original technique for conversion of H_2S to sulfur was the Claus-Chance Process. It has been modified considerably in recent years. In fact the Mathieson Chemical Company has developed a considerably improved process which has been used successfully in quite a few modern installations. A flow diagram of the modified Claus-Chance cycle used in the Mathieson process is shown in Figure 2. 1. 1. -30.

The first step consists of burning the feed gas in a specially designed reactor furnace. Flue gases from this furnace are partially cooled in a waste-heat boiler and then run to a catalyst converter. After passing through the first stage of the converter, the gases are run to a boiler feed-

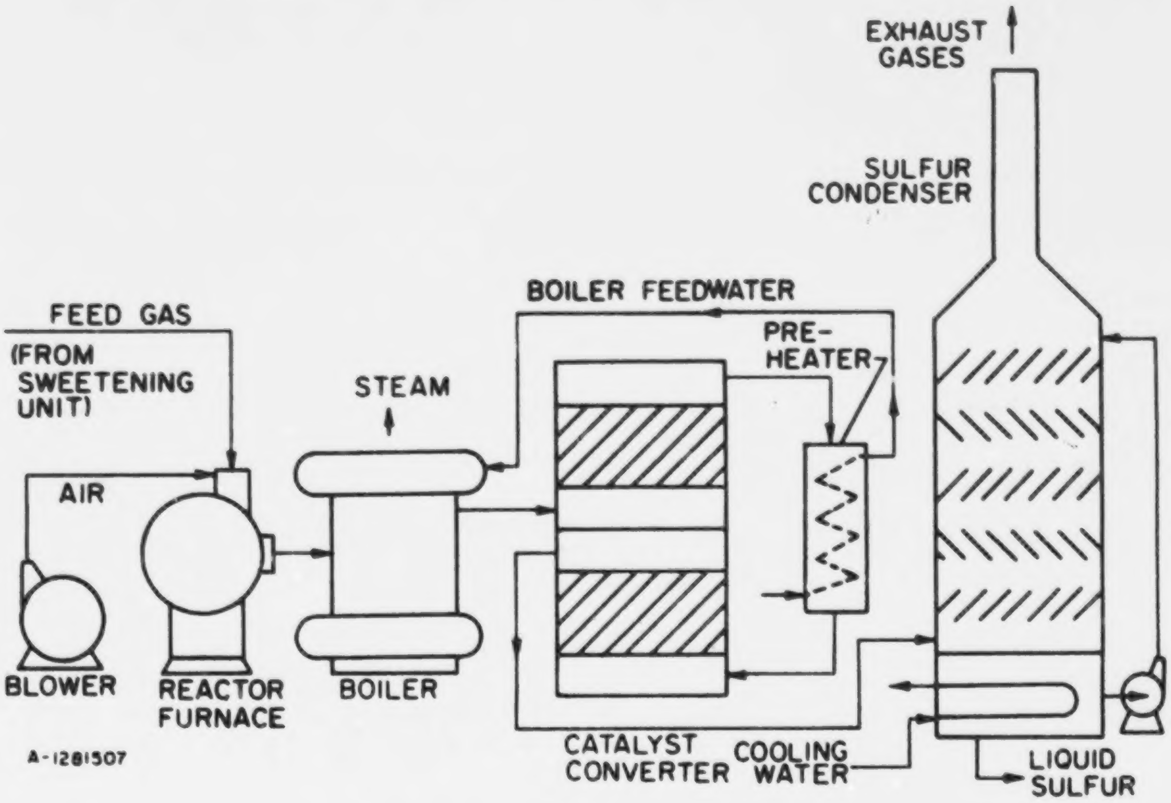

EXHAUST GASES
SULFUR CONDENSER
BOILER FEEDWATER
FEED GAS
(FROM SWEETENING UNIT)
STEAM
PRE-HEATER
AIR
BLOWER
REACTOR FURNACE
BOILER
A-1281507
CATALYST CONVERTER
COOLING WATER
LIQUID SULFUR

Figure 2.1.1.-30.SCHEMATIC FLOW DIAGRAM OF MODIFIED CLAUS-CHANCE (Mathieson) PROCESS

water economizer and then back through the second stage of the converter.

The final step is to pass the gases into a wash tower where they are cooled and make sulfur is condensed by direct contact with a recirculated stream of liquid sulfur. Exit gases from the tower, vented to the atmosphere, are primarily nitrogen, carbon dioxide, and water vapor.

In the reactor furnace, which is a refractory-lined horizontal cylindrical vessel, the feed gas is burned with a controlled amount of air. The air/gas ratio is carefully regulated so that one-third of the hydrogen sulfide in the acid gas is oxidized to sulfur dioxide and all hydrocarbons in the feed gas are completely burned.

In addition to the principal SO_2-producing reaction, two secondary ones may occur: the direct combustion of H_2S to sulfur and the reaction of SO_2 with unburned H_2S to form elemental sulfur.

The principal reaction taking place in the furnace — conversion of H_2S to SO_2 — is highly exothermic. Heat from this reaction, plus that from the combustion of hydrocarbons, results in the gas leaving the furnace at a temperature of as high as 1900°F. This temperature is well above that desired for optimum conversion of SO_2 to sulfur in the first stage of the catalytic reactor; hence excess heat is used to produce steam. Steam is produced in the waste-heat boiler, cooling the flue gases to approximately 650°F. This is generally sufficient steam for driving the air blower and sulfur reflux pump, for steam jacketing all sulfur flow lines, and for heating the coils in the run-down pit.

A bypass is provided around the boiler to permit passage of combustion gases from the furnace directly into the first converter. This provides satisfactory control of the converter's operating temperature. Such control is important because, if the temperature is too low, liquid sulfur may condense on the catalyst bed and in the flow lines and, if it is too high, recovery of sulfur is reduced, as pointed out earlier.

The catalytic converter contains two beds of natural bauxite catalyst, separated from one another by a diaphragm in the center of the vessel. The combined gas streams from the boiler and the bypasses are introduced into the first-stage bed where a considerable portion of the sulfur is formed.

The principal reaction taking place (H_2S reacting with SO_2 to form sulfur) is exothermic; consequently, the gas stream must be cooled before passing to the second-stage converter. Accordingly, the stream is passed to the boiler feedwater economizer for cooling to the desired temperature. A bypass around the economizer permits temperature control of the gases fed to the second-stage bed. Optimum operating temperature of this bed is about 500°F, a temperature which results in maximum sulfur conversion without the deposition of liquid sulfur on the catalyst.

From the second bed the gases enter the bottom of the sulfur wash tower. The rising gases contact cooled sulfur (liquid), which cools and condenses sulfur vapors contained in the gases. Liquid sulfur falling to the bottom of the wash tower is cooled by water-cooling coils in the bottom of the tower.

Circulation of cooled sulfur over the tower is provided by a steam-driven, submerged, cast iron centrifugal pump. As the vapor pressure of sulfur is extremely low at the gas outlet temperature, essentially all elemental sulfur is stripped from the rising gases. Sulfur in excess of recycling requirements is drawn off from the bottom of the tower and flows to a run-down pit.

This pit, if sufficient size to hold 2 day's production, is equipped with submerged steam coils to keep the sulfur in a molten state. Once a day sulfur is pumped to large vats where it is allowed to cool and solidify. In some instances, however, the liquid sulfur is pumped directly to steam-coil-equipped railroad tank cars for shipping in the molten state.

Operation of the plant is extremely simple and highly automatic. Air and gas feed to the furnace are regulated by a proportional ratio controller. All other equipment is provided with automatic controls so that once on-stream a minimum of manual operation is required.

Principal process control is maintained by analyzing gas from the final converter, using a modified Tutweiler analysis. An initial titration with standard iodine solution indicates the total quantity of H_2S plus SO_2 present in the gas; a subsequent titration of the resulting solution with standard acid gives an indication of its relative quantities.

Steam generated in the waste-heat boiler is utilized for driving the air blower and the sulfur-circulating pump. A recording pressure regulator in this system automatically bleeds steam into the system used for heating the run-down pit and all jacketed liquid-sulfur lines.

In addition to these steam-jacketing provisions, auxiliary gas burners are also provided for heating the major pieces of equipment in the event sulfur solidification does occur. After a prolonged shutdown, the auxiliary gas burners are installed through special ports provided for this purpose and are used to bring the furnace and other pieces of equipment to operating temperature. All sulfur lines are designed so that they drain completely in the event of a shutdown.

The Shamrock Oil and Gas Company's McKee plant at Sunray, Texas, is typical of this design. This plant recovers about 30 tons of elemental sulfur daily, using as feed gas the off-gas from a Girbotol(amines) gas treating unit, which processes about 340 million CF of sour gas each day. The sulfur plant feed gas totals about 1.9 million CF/day and enters the plant at about 100°F and 3 psig. Average concentrations are hydrogen sulfide, 43%; carbon dioxide, 55.5%; and light hydrocarbons, 1.5%.

The Claus-Chance Process was developed in 1890 and involved the vapor oxidation of H_2S to sulfur in the presence of a catalyst such as bauxite. The reaction was characterized by the evolution of large quantities of heat.

$$6\ H_2S + 3O_2 \rightleftarrows 6\ H_2O + 6\ S \qquad (1)$$

Investigation showed that the lower the operating temperature the greater the yield that could be obtained. Process problems consisted mostly of design for lower temperature operation.

Modification of the Claus-Chance Process was the result of developments by German industry during the past 75 years. By utilizing the fact that the overall reaction will proceed in two discrete steps

$$2\ H_2S + 3\ O_2 \rightleftarrows 2SO_2 + 2H_2O$$
$$\Delta H = 223{,}000\ \text{Btu/lb-mole}\ H_2S \qquad (2)$$

$$2\ H_2S + SO_2 \rightleftarrows 3/8\ S_8 + 2H_2O$$
$$\Delta H = 44{,}800\ \text{Btu/lb-mole}\ SO_2 \qquad (3)$$

the problem of low yield because of high temperature in the presence of the catalyst is avoided.

This is accomplished by first burning a portion of the H_2S to SO_2 (Reaction 2) and then reacting the remaining H_2S with the SO_2 in the presence of the catalyst to yield sulfur and water (Reaction 3).

Almost 80% of the total heat of reaction will be evolved before the catalytic oxidation reaction takes place. Hence, the temperature rise in the catalytic vessel will be much less and the yield much greater.

Also, a very high-temperature noncatalytic combustion of H_2S with air in stoichiometric or greater proportion will produce sulfur directly (Reaction 1) in yields of about 60%.

This high-temperature combustion can be followed by one or more Claus-Chance catalytic converters to recover nearly all the sulfur.

Compared with the original Claus-Chance Process these modifications were capable of higher sulfur yields and the plant size required for equivalent sulfur production was substantially reduced.

As a result of the German development work, two basic schemes are used today. These are the once-through and split-stream processes. Variations of each exist to handle differences in acid-gas quantity, composition, and sulfur recovery level.

A schematic diagram of the once-through Claus-Chance Process is shown in Figure 2.1.1.-31. In this process, sufficient air to convert all the H_2S to sulfur is added to the acid-gas stream, and the gases enter the acid-gas burner and reaction chamber.

The H_2S and oxygen react catalytically, as discussed for Reaction 1 above, to produce sulfur vapor. Temperatures here will be close to 2000°F. As much as 50-60% conversion of the hydrogen sulfide to sulfur may take place in this reaction chamber. Some SO_2 will also be formed. Most of the heat generated in the burner and reaction chamber is removed in the waste-heat boiler by generating high-pressure steam. The gas is cooled to about 600°-700°F. In the sulfur condenser the gases are cooled further to about 300°F and the sulfur is condensed and drained to storage.

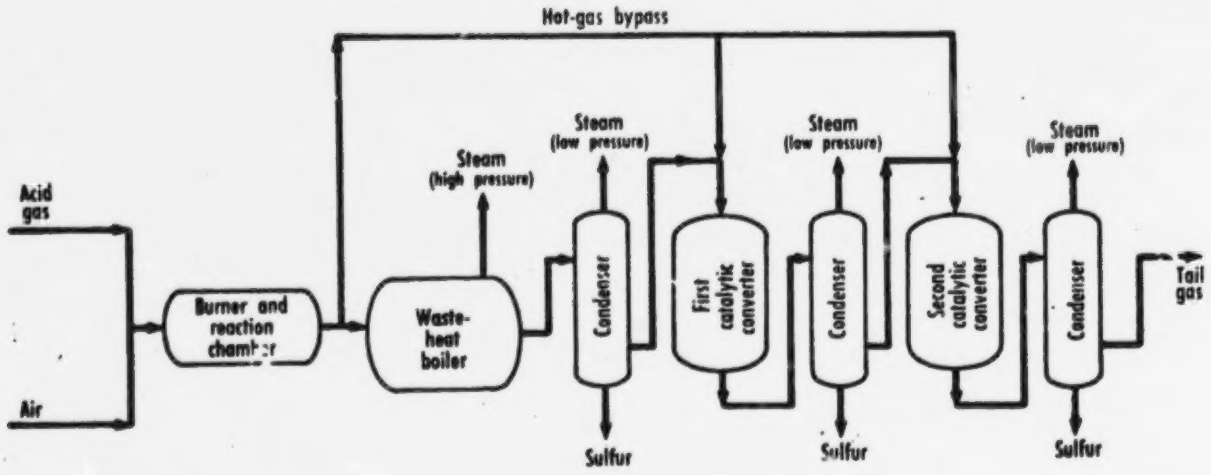

Hot-gas bypass
Acid gas
Air
Burner and reaction chamber
Steam (high pressure)
Waste-heat boiler
Steam (low pressure)
Condenser
Sulfur
First catalytic converter
Steam (low pressure)
Condenser
Sulfur
Second catalytic converter
Steam (low pressure)
Condenser
Sulfur
Tail gas

Figure 2.1.1.-31. ONCE-THROUGH CLAUS-CHANCE PLANT[52]

The condenser also generates low-pressure steam.

The gases are reheated by a hot-gas bypass and introduced to the first catalytic converter at about 450°-500°F. This procedure allows the sulfur produced in the burner to be removed and the temperature of the gases to the catalyst bed to be optimally controlled for high yield.

From the first catalytic converter the effluent gas is cooled to about 300°F in a condenser where more low-pressure steam is generated. Liquid sulfur is drained to storage and the remaining acid gases are reheated (450°-500°F) and passed through the second catalytic converter, followed by a sulfur condenser. Liquid sulfur is removed and low-pressure steam is generated.

The remaining acid gases leave the process by an incinerator where any remaining sulfur and H_2S are burned to SO_2 before venting to the atmosphere. Two heat-exchange methods can be used for reheating the acid gases from the sulfur condensers.

A hot-gas bypass around the waste-heat boiler can be used as shown in Figure 2.1.1.-31, or an in-line acid-gas burner employing a small flow of the acid-gas and air mixture can be used for direct-fired heating.

One variation of this process consists of omitting the sulfur condenser between the waste-heat boiler and the first catalytic converter. Costs are less, but yield is reduced.

The split-stream Claus-Chance Process is shown schematically in Figure 2.1.1.-32. In this process one-third of the acid-gas stream and sufficient air to burn the H_2S to SO_2 are fed to the sulfur burner-waste heat boiler vessel. Very little sulfur will be produced here, as most of the H_2S is burned completely to SO_2.

The heat of reaction evolved is used to generate high-pressure steam. The effluent gases are then mixed with the other two-thirds of the acid-gas stream and passed through the first catalytic converter at about 450°-500°F. The remaining H_2S and the SO_2 react (Reaction 3) to form sulfur vapor and water.

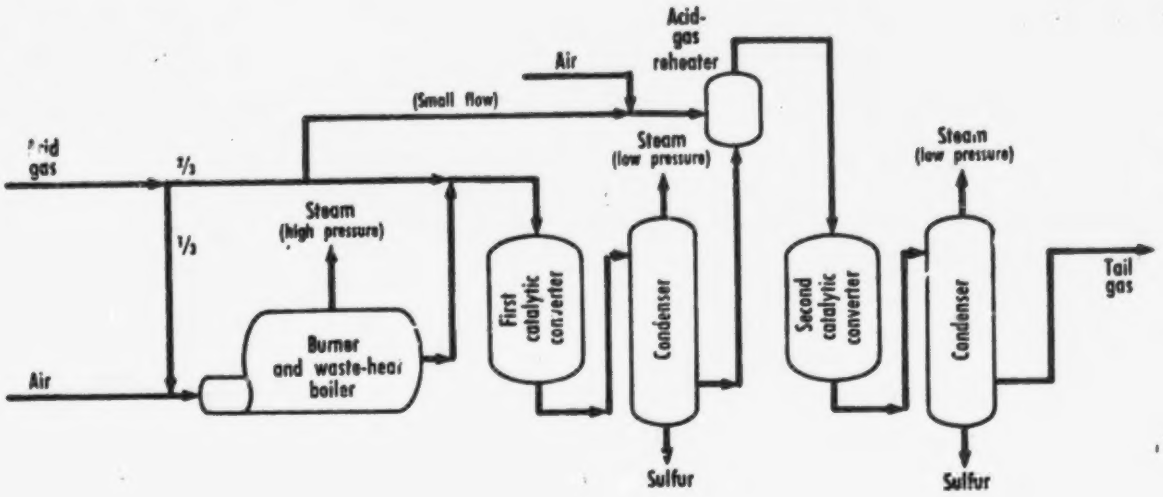

Acid-gas reheater
Air
(Small flow)
Acid gas
2/3
1/3
Steam (high pressure)
Steam (low pressure)
Steam (low pressure)
Burner and waste-heat boiler
Air
First catalytic converter
Condenser
Second catalytic converter
Condenser
Tail gas
Sulfur
Sulfur

Figure 2.1.1.-32. SPLIT-FLOW CLAUS-CHANCE PLANT[52]

The gases then pass to a condenser where the sulfur is condensed at about 300°F and is withdrawn as liquid. The remainder of the stream is reheated to about 450°F and passed through a second catalytic converter and a sulfur condenser. The effluent gases are then sent to an incinerator, and the combustion gases are vented to the atmosphere. Low-pressure steam is generated in the sulfur condensers.

An in-line acid-gas-fired reheater is shown, but heat exchange with the hot acid-gas stream leaving the first catalytic reactor could be employed.

The direct-oxidation process has been developed by Pan American Petroleum Corp. to process acid-gas streams of low-to-medium H_2S concentrations and relatively large amounts of hydrocarbons. Grekel[27] has reported on the development work. The process is based on the principle of direct oxidation of H_2S to sulfur over bauxite catalysts:

$$2\ H_2S + O_2 \rightleftharpoons 2\ S + 2H_2O \qquad (4)$$

High space velocities in an excess of air are employed, and sulfur recoveries are independent of the Claus-Chance equilibrium and temperature effects.

For example, where Claus-Chance equilibrium calculations would indicate about 50% conversion to sulfur, actual recoveries of about 75% can be obtained.

In the presence of paraffin hydrocarbons, H_2S is oxidized to sulfur without the simultaneous oxidation of the hydrocarbons as the temperature limits in the reactors are held to less than 1000°F.

A schematic diagram of the process is shown in Figure 2.1.1.-33. Air is added to the preheated feed stream and then passed through the catalyst bed in the oxidation reactor. Products include sulfur vapor and some SO_2. This stream is passed through a sulfur condenser, steam is generated, and the gases are fed to a second stage of catalytic conversion followed by sulfur condensation. Additional air is added prior to the second catalytic stage. Sulfur recoveries of 80-85% can be expected.

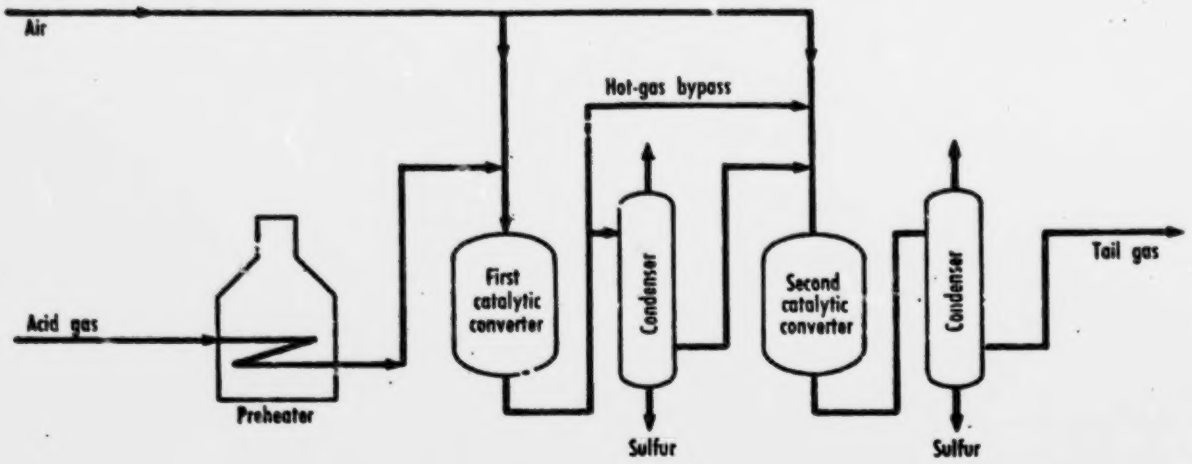


Figure 2.1.1.-33. DIRECT OXIDATION PROCESS[52]

The process is applicable to gas streams that cannot be economically processed by any other method. Goar[26] reports on the application to acid-gas streams of 7-10% H_2S, 70-90% CO_2, and some hydrocarbons. High-quality sulfur is produced in quantities of 4-15 long tons/day.

2.1.1.2.9.2. Investment Costs

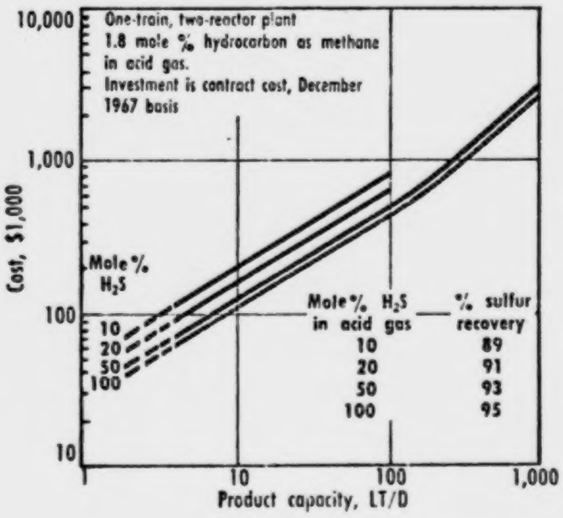
Investment costs for sulfur recovery plants have been decreasing as new technology has developed. A major contributor to this trend has been the use of package-plant units, as originated in 1956 by Pan American Petroleum Corp. This is illustrated in Table 2.1.1.-6, which compares the actual cost of a plant built in 1952 with the present cost (1967) for an equivalent plant.

The investment cost of sulfur plants is a function of plant capacity and concentration of hydrogen sulfide in the feed gas. This is shown in Figure 2.1.1.-34, which represents the 1967 contract price for one-train, two-reactor plant, either straight through or with a split-train. Individual plants may vary ± 20% from these typical curves, depending on specific conditions. Also, as shown in the Figure 2.1.1.-34, a plant to process a 20% hydrogen sulfide stream will cost about 40% more than a plant to process a 100% hydrogen sulfide stream, for the same tonnage. This is primarily because of the larger vessels required for the dilute streams for equivalent sulfur capacity.

Table 2.1.1.-6. INVESTMENT COSTS FOR SULFUR RECOVERY UNITS[28]

Production Capacity, long tons/day	18	52
Acid-Gas Analysis, mole % H_2S (dry basis)	62	17
Actual Contract Cost,* 1952, $	260,000	491,000
Contract Cost Escalated to 1967 Construction Cost Index (1.75 X 1952 cost), $	450,000	860,000
Contract Cost in 1967 for Present Design, $	160,000	430,000
Percent of 1952 Plant Cost	62	87
Percent of 1967 Cost for 1952 Plant Design	36	50

* Both plants had two reactors. Contract cost does not include land, inventory and working capital, owner's overhead and engineering, boiler feed water treating, steam condensation, or incinerator and stack.



One-train, two-reactor plant
1.8 mole % hydrocarbon as methane in acid gas.
Investment is contract cost, December 1967 basis

Figure 2.1.1.-34. INVESTMENT FOR CLAUS-CHANCE SULFUR RECOVERY PLANT[28]

The average sulfur recovery plant capacity in the U.S. has not changed in recent years. In 1961, the average was 54 long tons/day for natural gas plants, and 62 long tons/day for plants in refineries. Since 1961 additions have added 54 long tons/day and new plants have added 52.

However, the maximum size of sulfur recovery plants has increased markedly. Before 1961, the largest plant was Texas Gulf Sulfur's Worland, Wyoming, unit, which had a 400 long tons/day capacity. This unit operated from 1950 through 1967, when it was finally shut down. The largest today is Petrogas' Balzac, Alberta, plant, which has a capacity of 1970 long tons/day. By the end of 1968 there will be four plants of over 1000 long tons/day capacity each in operation in Canada.

2.1.1.2.9.3. Operating Costs

Operating costs include labor (both for operating and maintenance), catalyst, sulfur loading, power for air blowers, and the fuel and operating costs of incinerator and disposal stacks.

The power required by the air blower depends on the hydrogen sulfide concentration in the gas, the concentration of hydrocarbons (especially any heavy hydrocarbons), the number of reactors, and the design pressure drop. Cost will also depend on whether a steam turbine or electric drive is used. An average value to use for preliminary estimates is $0.35/long ton, assuming power required is about 2 hp/long ton-day of capacity.

Experience indicates that even plants with up to several hundred tons per day of capacity do not require full-time operation. These are generally handled as a part of the sweetening unit operation, with occasional tests of the residue gas and checks of the controls and recorders.

Loading operations are generally done by contract labor. For tank cars loading liquid sulfur, the costs will be $0.10-$0.15/long ton loaded. For gondola cars loading solid sulfur, costs vary from $0.55 to $0.65/long ton loaded and will be higher when crushing is required.

The costs for operating and maintenance labor, and catalyst are shown in Figure 2.1.1.-35 as a function of plant capacity.

The effect of plant size and feed gas composition on the total cost of sulfur recovery from acid gas is shown in Figure 2.1.1.-36. This is based on the assumption of constant income for 20 years, an income tax rate of 50%, and a 7% investment tax credit. It is also assumed that a 12% rate of return (after tax) on 100% equity investment is obtained if the annual net income before tax is 20% at the investment. These figures would have to be

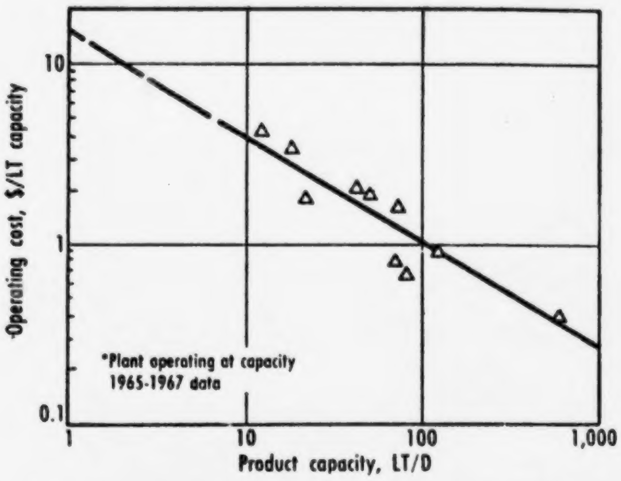


Figure 2.1.1.-35. OPERATING COST OF CLAUS-CHANCE SULFUR PLANT[28]

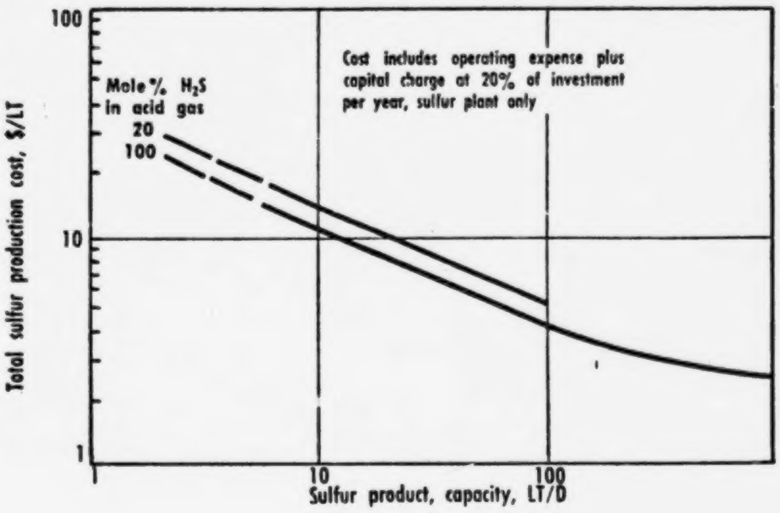


Figure 2.1.1.-36. TOTAL COST OF SULFUR RECOVERY FROM ACID GAS BY CLAUS-CHANCE PROCESS[28]

revised if a declining production operation is considered. Also, since investment and operating expenses vary from area to area, these figures should be considered only for first estimates.

2.1.1.2.10. Other H_2O/CO_2 Purification Systems

2.1.1.2.10.1. Process Description

2.1.1.2.10.1.1. Seaboard Process

In the Seaboard Process a dilute solution of sodium carbonate (3%) is used as an absorbing medium. Solution regeneration is accomplished by blowing air through the spent solution in a separate tower. The process is diagramed in Figure 2.1.1.-37.

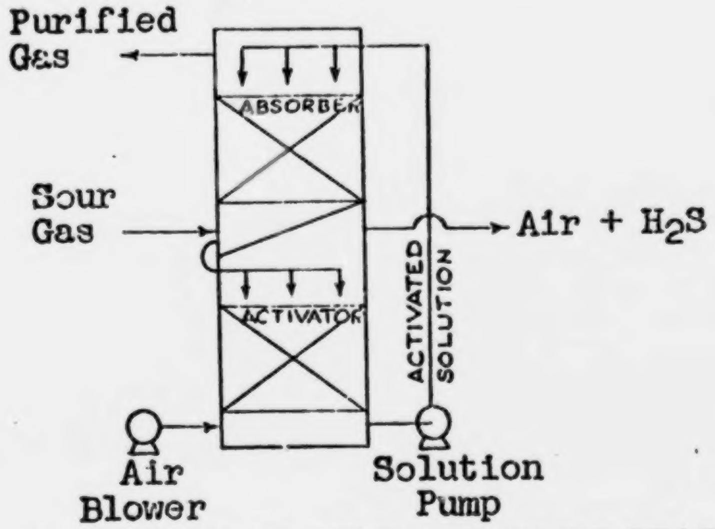

Purified Gas
ABSORBER
Sour Gas
Air + H_2S
ACTIVATOR
ACTIVATED SOLUTION
Air Blower
Solution Pump

Figure 2.1.1.-37. SCHEMATIC FLOW DIAGRAM OF SEABOARD PROCESS

The inexpensive reagent and the small space requirements made this process attractive in the past. However, since it is relatively inefficient in sulfur removal, it has now largely been replaced by amine scrubbing.

This system is very similar to the iron-soda process and uses a dilute solution of sodium carbonate to absorb the H_2S. Best operation is obtained with solutions containing about 3% Na_2CO_3. Regeneration is obtained by aeration in a separate tower; the H_2S is either vented to the atmosphere with the regenerative air or may be used as furnace feed if it constitutes a menace.

Absorption efficiencies of up to 90% of the H_2S in the feed gas can be obtained at low pressures; for higher pressures, the efficiency increases rapidly. Some of the CO_2 and almost all HCN, if present, will be absorbed. However, the process is unsatisfactory for large quantities of HCN or certain other contaminants due to side reactions that occur. The principal advantage of the process is its simplicity, but, because of the relatively low H_2S removal efficiency and high operating costs, its use is not expanding.

2.1.1.2.10.1.2. Iron-Soda Process

This process represents an attempt to utilize the chemical reaction of the iron oxide process in a continuous system. The basic cycle is shown schematically in Figure 2.1.1.-38. In an absorber the sour gas contacts a dilute solution of sodium carbonate containing suspended iron oxide or hydroxide. The spent solution is regenerated in a horizontal aerator by bubbling air through the solution. The free sulfur formed rises to the top of the liquid and is skimmed off, and the regenerated solution returns to the top of the contactor.

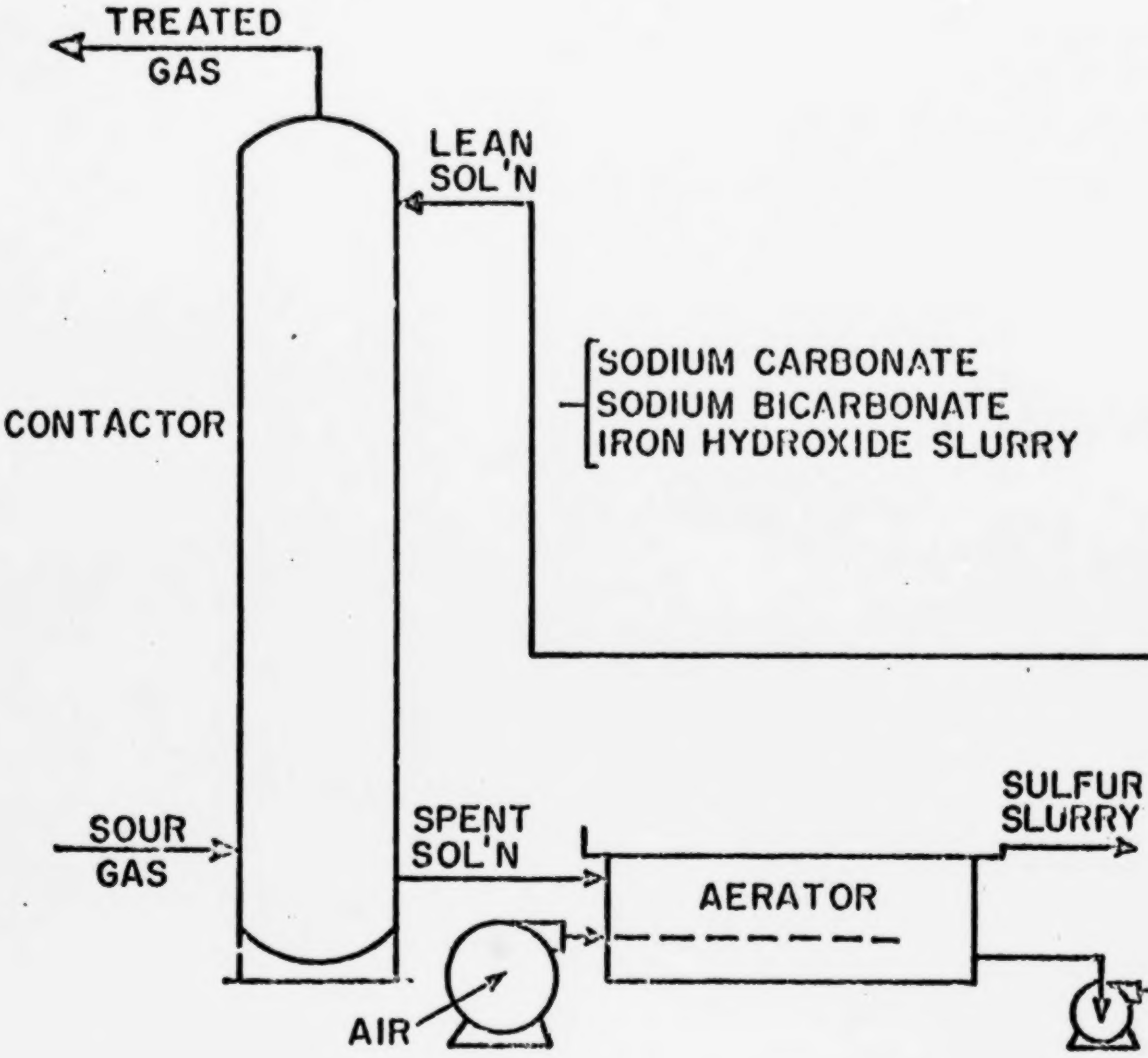

TREATED GAS
LEAN SOL'N
CONTACTOR
SODIUM CARBONATE
SODIUM BICARBONATE
IRON HYDROXIDE SLURRY
SOUR GAS
SPENT SOL'N
SULFUR SLURRY
AERATOR
AIR

Figure 2.1.1.-38. SCHEMATIC FLOW DIAGRAM OF BASIC IRON-SODA PROCESS

Two reactions take place in the contactor:

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

$$3NaHS + 2Fe(OH)_3 \rightarrow 2FeS + S + 3NaOH + 3H_2O$$

Recent experimental data on this type of process indicate that the optimum solution concentrations are approximately 0.4% $Fe(OH)_3$, 0.6% Na_2CO_3, and 3.0% alkali (Na_2CO_3 + $NaHCO_3$). Removal of 88-95% of the H_2S initially present in the sour feed can be obtained under these conditions when the regeneration reaction is —

$$4FeS + 3O_2 + 6H_2O \rightarrow 4Fe(OH)_3 + 4S$$

which indicates that considerable quantities of air are required to reactivate the solution. Recovery of from 70 to 85% of the sulfur absorbed from the sour gas can be obtained, the remainder forming nonregenerable sulfur salts. The product contains about 50% water and 20% ash, which is principally solid $Fe(OH)_3$ trapped in the free sulfur.

Recently this process has been applied to the removal of small quantities of H_2S from natural gas produced in the Santa Maria oil field of California. A selective process for removing H_2S was required as the gas contains about 15% CO_2 and it was desired to remove most of this CO_2 for the production of dry ice by monoethanolamine in a later process. The iron-soda plant reduced the H_2S content from 34 to 3 or 4 grains of H_2S per 100 SCF of gas without appreciably affecting the CO_2 content.

2.1.1.2.10.1.3. Thylox Process

A schematic diagram of the Thylox Process is shown in Figure 2.1.1.-39. The solution of various sodium and ammonium thioarsenates combines in the absorber with H_2S to form compounds containing more sulfur. The reaction for the absorption of H_2S is –

$$Na_4As_2S_5O_2 + H_2S \rightleftharpoons Na_4As_2S_6O + H_2O$$

Very little CO_2 is absorbed because of the low alkaline nature of the solution, but HCN is completely absorbed and generates thiocyanates. H_2S removal efficiencies of 95-99.7% have been obtained.

Regeneration of the spent solution is obtained by having the solution contact air in a vertical aerator or pressure thionizer. The regeneration reaction is –

$$Na_4As_2S_6O + 1/2\ O_2 \rightleftharpoons Na_4As_2S_5O_2 + S$$

The elemental sulfur formed can be recovered as a marketable byproduct. Arsenic and Na_2CO_3 are added to the spent solution stream as makeup, and thionized during regeneration.

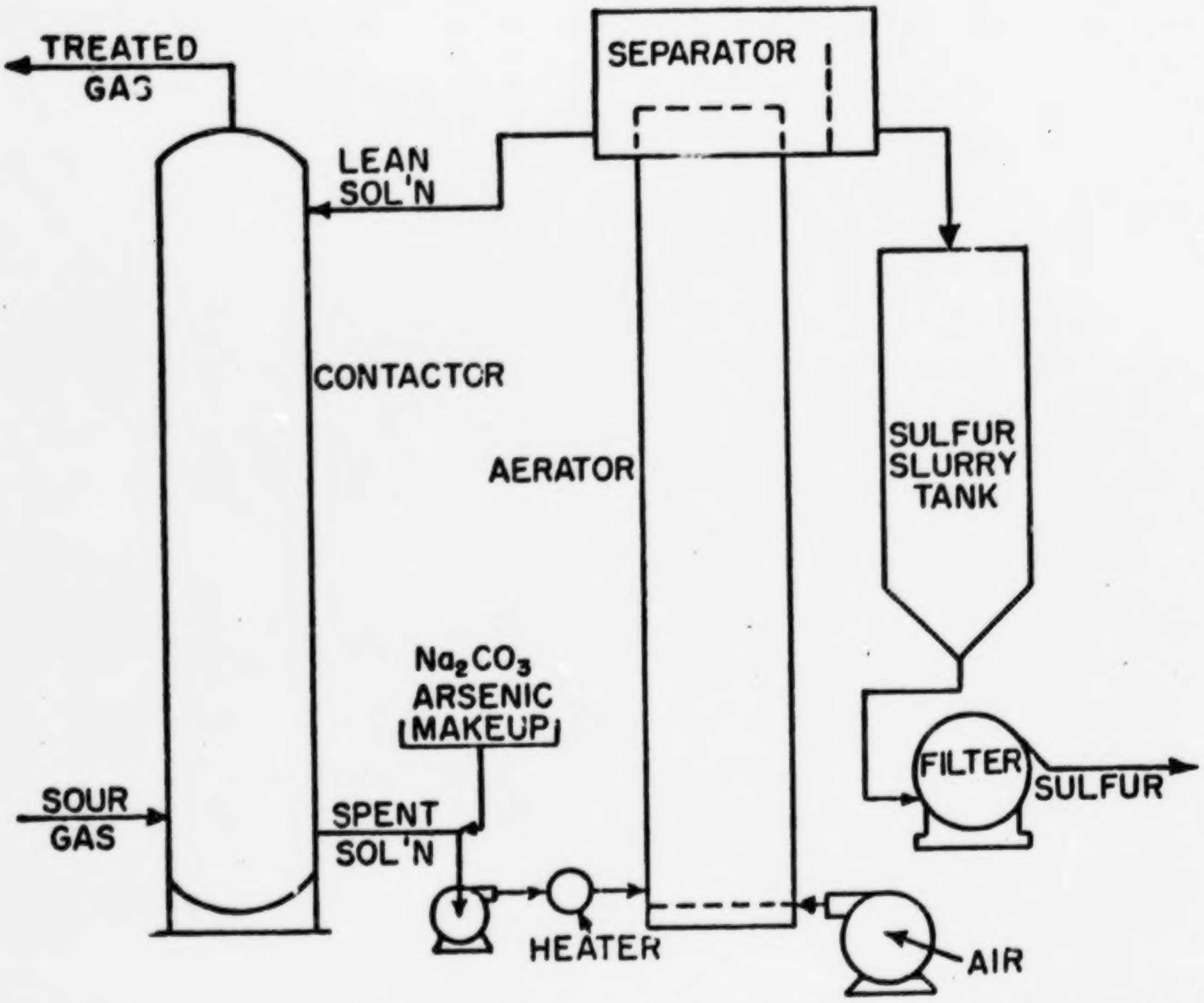

TREATED GAS
SEPARATOR
LEAN SOL'N
CONTACTOR
AERATOR
SULFUR SLURRY TANK
Na_2CO_3 ARSENIC MAKEUP
SOUR GAS
SPENT SOL'N
FILTER
SULFUR
HEATER
AIR

Figure 2.1.1.-39. SCHEMATIC FLOW DIAGRAM OF THYLOX PROCESS

If no credit is taken for the by-product sulfur recovered by this process, the operating costs are relatively high. The process is also more complex than some of the others. However, H_2S removal efficiency is very good and is selective in the presence of CO_2.

2. 1. 1. 2. 10. 1. 4. Fluor Solvent Treating

In 1960, three patents were issued to Fluor Corp.[41, 40, 59] covering the organic solvents propylene carbonate, glycerol triacetate, butoxy-diethylene-glycol acetate, and methoxy-triethylene-glycol acetate when used for the adsorption of carbon dioxide from high-pressure gas streams.

All of these solvents resulted from a research program aimed at developing a sweetening agent that would simply absorb the acid gases by physical solution (such as a water wash) and permit solvent regeneration by release of pressure and without heat application.

Although originally developed for removing large concentrations of CO_2 from natural gas, use of these solvents has been proposed for CO_2-H_2S removal from high-pressure natural gas streams, for CO_2 removal from synthesis-gas streams, and for selective removal of H_2S from gas that contains both CO_2 and H_2S.

Operation of the process depends on high-pressure gas with high-acid-gas concentration because the partial pressure of the acid gases is the driving force toward the solution. Since only a physical absorption of the gases is involved, the carrying capacity of the solvent will be almost directly proportional to the partial pressure of the acid gases.

This is a very important consideration because the circulation rate required to treat a given volume of gas will be nearly independent of the acid-gas content of the sour gas. With amine or hot potassium carbonate processes involving absorption with chemical reaction, the circulation of the sweetening agent is governed by the acid-gas content and the quantity of acid gas to be removed.

The solvents will pick up water vapor, carbonyl sulfide, carbon disulfide, mercaptan, and hydrocarbons as well as CO_2 and H_2S. H_2S is several times more soluble than CO_2. On this basis, a selective H_2S-removal process could be operated. The dissolved hydrocarbons — particularly propane, butane, and natural gasoline — may be lost with the acid gases unless specific recovery steps are taken. The solvents are noncorrosive to carbon steel. Low solvent vapor pressures result in low vaporization losses. Chemical stability is good with no solvent degradation experienced.

A typical application of the Fluor Solvent Process, as discussed by Kohl and Buckingham,[39] is shown in Figure 2.1.1.-40. Sour gas is contacted countercurrently with the solvent in a conventional absorption column. The lean solvent enters the top of the absorber at ambient temperature or below, and the spent solution leaves the bottom at a slightly higher temperature due to the heat of absorption of the acid gases.

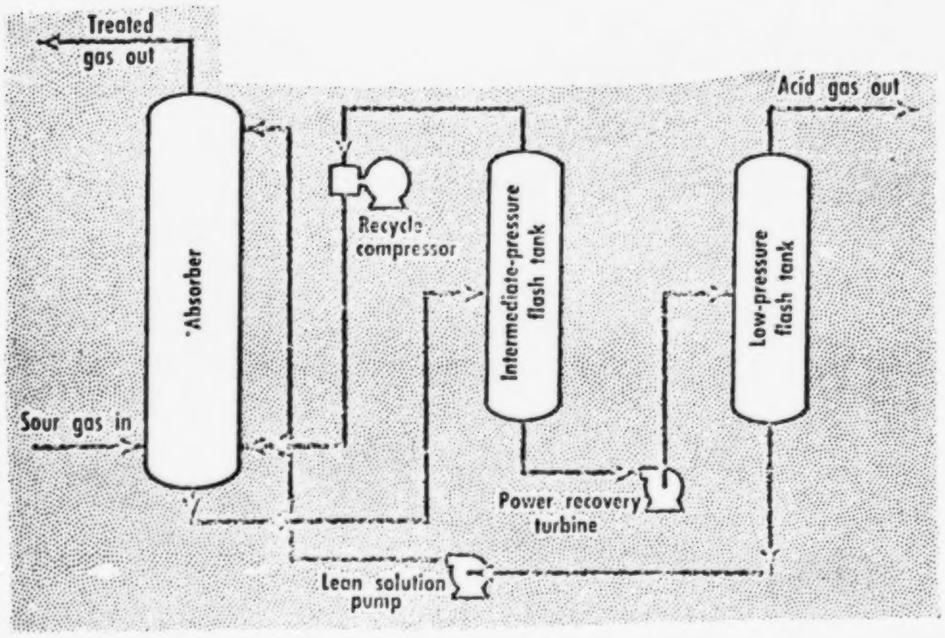

Treated gas out
Acid gas out
Absorber
Recycle compressor
Intermediate-pressure flash tank
Low-pressure flash tank
Sour gas in
Power recovery turbine
Lean solution pump

Figure 2.1.1.-40. FLUOR SOLVENT TREATING[50]

In addition to CO_2, H_2S, and other sulfur-containing compounds, the solvent will absorb water vapor and a small quantity of hydrocarbons from the gas stream.

The spent solution from the absorber goes through two stages of flashing to disengage the absorbed gases and regenerate the solvent.

The intermediate-pressure flash tank is operated at a relatively high pressure, but releases a major portion of the dissolved hydrocarbons. The vapor from this stage is compressed to absorber pressure and recycled back to the bottom of the absorber.

The solvent from this stage can be passed through a power recovery turbine to the atmospheric-pressure flash tank for the dissolution of the remaining acid gases, which are vented to the atmosphere.

The temperature of the solution will decrease during the flash steps because of the heat required by acid-gas desorption and expansion. The regenerated cool solution is recycled to the top of the absorber.

With this process scheme, the regeneration step is completed with the atmospheric flash. The lean solution returned to the absorber contains acid gases still in solution equivalent to an atmospheric partial pressure; the acid-gas content of the sweetened gas could not be reduced below this level.

For example, if the gas stream pressure was 800 psig and the acid gas was essentially CO_2, the treated gas could not be reduced to a CO_2 level below 2%.

Variations of this scheme can be made depending on the temperature and composition of the sour-gas stream, and the degree of acid-gas removal desired. If a low acid-gas content is desired, air stripping and/or vacuum flashing of the solution may be used with the solvent following the atmospheric flash stage.

If the gas flashed from the intermediate flash stage is rich in hydrocarbons, the required solution flow rate can be reduced — at the expense of absorber height — by feeding the recycle gas into the absorber at a point below the main gas inlet, or the intermediate-pressure flash gas may be made to contact a separate stream of solvent before recycle and then be fed back into the absorber at a point above the main gas inlet. This reduces the size of both the recycle compressor and the main absorber.

Should subambient-temperature operation be desired, an appropriate heat integration scheme will provide most of the necessary heat removal. For very high pressure gas streams, economics may dictate the use of several flash stages and several power recovery turbines.

Selective H_2S removal has been proposed by sharply reducing the solution circulation rate through the absorber. The selectivity would be obtained because H_2S is several times more soluble than CO_2.

Buckingham[6] has reviewed the operation of some of the existing Fluor solvent plants. All but one of these plants use the propylene carbonate solvent.

The largest unit processes 220 million SCF/day of natural gas containing 53% CO_2 at 850-1000 psig. The residue gas leaves with 2% CO_2, and 3 gr/100 SCF of H_2S is reduced to the pipeline specification of 0.25 gr/100 SCF. The solvent (propylene carbonate) also dehydrates the gas to pipeline specifications and substantially lowers the mercaptan content.

Another plant is designed for an absorption pressure of 1000 psig and a widely varying acid-gas content of 10-30% CO_2 and 5-15% H_2S. Here, sulfur recovery with a Claus-Chance Process is used. One plant employs subambient-temperature operation. Of the hydrocarbons absorbed by the propylene carbonate, 98% are recovered.

The simplicity of the process results in low investment and operating costs when applied to high-pressure, high-acid-gas-content gas streams. The process bears investigation for gas streams with at least 60-100 psia acid-gas partial pressure or at least 25 psia H_2S partial pressure, if selective H_2S removal is contemplated. Economics will be less attractive if the natural gas stream is "wet" or if high-purity treated gas is required.

2.1.1.2.10.1.5. Water-Wash Process

A water wash is considered for bulk scrubbing of sour gas, but it must be followed by a second sweetening process to obtain a reasonable degree of sweetening.

The use of water washing for gas sweetening has two obvious advantages — low cost and ready availability of sweetening agent.

Its use for removal of bulk quantities of acid gas from high-pressure natural gas streams has been proposed by Burnham[8] and by Froning et al.[22] An excellent discussion is also presented in Kohl and Riesenfeld.[42]

Absorption of acid gases such as hydrogen sulfide and carbon dioxide in water is considered a physical absorption. The capacity of water to absorb acid gases is directly proportional to its partial pressure.

The process would be most effective with high-pressure gas with a high-acid-gas concentration. Since these acid gases form acids in solution, the possibility of corrosion must be considered. Regeneration of the water is by pressure reduction and flashing; stripping heat is not necessary.

Some other advantages of the water-wash process include:

- The solvent is inexpensive, and mechanical and vaporization losses are easily replaced.
- The solvent is nonreactive with carbonyl sulfide and is not degraded by oxygen.
- H_2S is several times more soluble than CO_2, and a water-wash scheme could be considered partially selective toward H_2S under certain conditions.

The principal disadvantage is that an additional process must be used to sweeten the gas to normal specifications. Other disadvantages of the process are 1) because of the low acid-gas solubilities, high pumping loads are required for water circulation; 2) hydrocarbon losses from solution in the water can be quite high; and 3) CO_2 removal efficiency is poor. In addition, corrosion will occur unless steps are taken for its reduction. Corrosion inhibitors, stress relieving of vessels, and the use of stainless steel for critical parts are necessary to minimize this corrosion.

A typical water-wash scheme for acid-gas removal from a high-pressure natural gas is shown in Figure 2.1.1.-41.

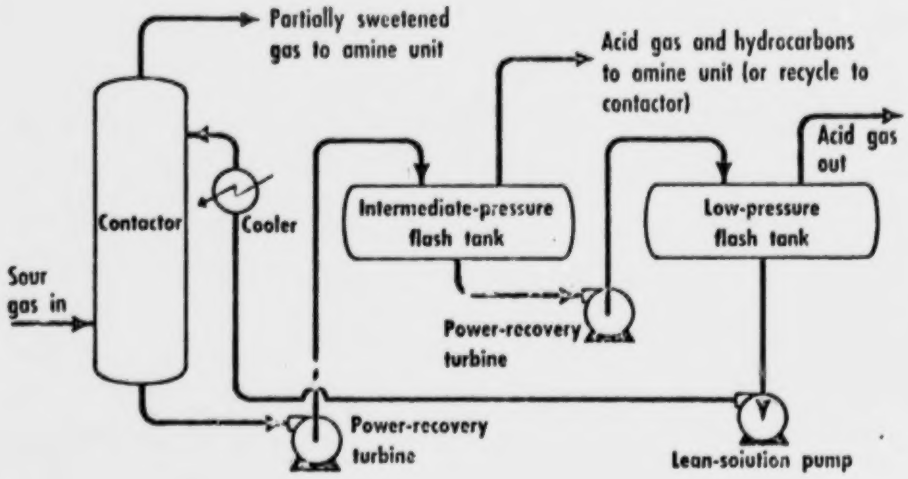

Partially sweetened gas to amine unit
Acid gas and hydrocarbons to amine unit (or recycle to contactor)
Acid gas out
Contactor
Cooler
Intermediate-pressure flash tank
Low-pressure flash tank
Sour gas in
Power-recovery turbine
Power-recovery turbine
Lean-solution pump

Figure 2.1.1.-41. WATER-WASH SWEETENING PROCESS[53]

The sour-gas stream countercurrently contacts the water in the absorber where the acid gases are picked up. The partially sweetened gas stream is sent to a second sweetening stage — possibly an amine unit if typical pipeline specifications must be met.

The water and acid-gas solution leaves the contactor, passes through a power recovery turbine, and is sent to an intermediate-pressure flash drum. Here some of the acid gases and hydrocarbons absorbed in the contactor are flashed from solution. These can be recycled to the contactor, sent to the amine (or other) sweetening unit downstream, or used for fuel. The purpose of the intermediate-pressure flash is to recover as much of the absorbed hydrocarbon as possible before the acid gases are flared or processed. The water stream from the intermediate-pressure flash can be passed again through a power recovery turbine and then to the low-pressure flash drum.

Even though the heat-of-solution effect will be very slight, a small cooler may be needed downstream of the low-pressure flash. The water is then pumped back to the top of the high-pressure contactor.

Burnham[8] reports on the operation of a commercial test unit and other studies of water washing high-pressure gas.

Results indicate that the efficiency of acid-gas removal is much better for high-H_2S-content gas. Water-wash sweetening with an amine cleanup reportedly has a 12-15% less investment and 50% less fuel cost than an equivalent amine unit for low H_2S/CO_2 ratio gas. For high H_2S/CO_2 gas, the combination will cost 40% less for investment and 70% less for fuel.

2. 1. 1. 2. 10. 1. 6. Vacuum Carbonate Process

The vacuum carbonate process is essentially a modification of the Seaboard Process. Regeneration of the spent sodium carbonate solution is by vacuum distillation rather than by air stripping.

Distillation of the spent carbonate solution allows the H_2S to be recovered in a concentrated, usable form.

Vacuum distillation of the solution results in a steam requirement of about one-sixth that required for steam stripping at atmospheric pressure.

A flow scheme for a vacuum carbonate plant is shown in Figure 2.1.1.-42. The absorber part of the process is as described for the Seaboard Process. The actifier is operated under vacuum – about 2.5 psia.

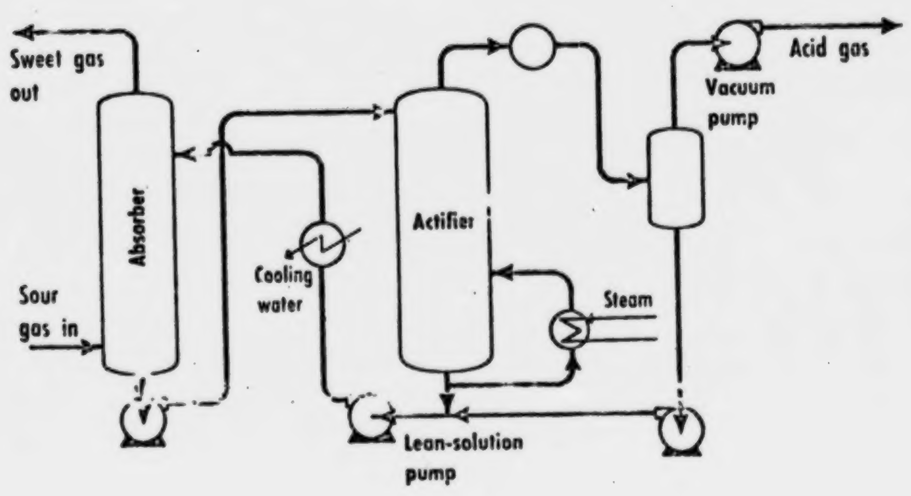

Sweet gas out
Absorber
Sour gas in
Cooling water
Actifier
Steam
Lean-solution pump
Vacuum pump
Acid gas

Figure 2.1.1.-42. VACUUM CARBONATE PROCESS[53]

The spent solution is pumped to the actifier and the regenerated solution is pumped from the bottom of the actifier through a solution cooler and back to the absorber. Gases, including water vapor, from the top of the actifier pass through a condenser and vacuum-pump system.

Operating data indicate that an 80-90% H_2S removal efficiency can be achieved. Other operating and design data as well as typical operating and investment costs are presented in Kohl and Riesenfeld.[42]

As with the Seaboard Process, the vacuum carbonate process has been mostly used for coke-oven-gas sweetening.

2. 1. 1. 2. 10.1. 7 Liquid Absorption-Oxidation Processes

Liquid absorption-oxidation processes are characterized by the Ferrox and Manchester Processes. In the 1920's the Ferrox Process was introduced by the Koppers Co. It has been used extensively both for natural gas sweetening and for coal-gas purification.

A British modification known as the Manchester Process employs a variation in flow scheme and is currently of considerable importance to the British gas industry. It is used to sweeten manufactured or town gas.

The chemistry of both of these processes is the same. The sweetening agent is iron oxide suspended in an alkaline, aqueous solution. Hydrogen sulfide reacts with sodium carbonate to give sodium hydrosulfide and sodium bicarbonate as in Reaction 1.

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3 \qquad (1)$$

Sodium hydrosulfide then reacts with iron oxide to form iron sulfide (Reaction 2).

$$Fe_2O_3 \cdot H_2O + 3NaHS + 3NaHCO_3 \rightarrow Fe_2S_3 \cdot 3H_2O + 3Na_2CO_3 + 3H_2O \qquad (2)$$

Regeneration consists of aeration of the spent solution — converting iron sulfide to elemental sulfur and iron oxide (Reaction 3).

$$2Fe_2S_3 \cdot 3H_2O + 3O_2 \rightarrow 2FeO_3 \cdot 3H_2O + 6S \qquad (3)$$

Besides these main reactions, which are the basis of the process, particularly troublesome side reactions can and often do occur:

$$2NaHS + 2O_2 \rightarrow Na_2S_2O_3 + H_2O \qquad (4)$$

$$Na_2S + 3/2\ O_2 + S \rightarrow Na_2S_2O_3 \qquad (5)$$

Reactions 4 and 5 show the formation of the thiosulfate salt. Hydrogen cyanide, if present, can also yield undesirable compounds. These nonregenerable salts can lead to high chemical replacement costs.

The Ferrox Process is shown schematically in Figure 2.1.1.-43. The sweetening solution is composed of about 3.0% sodium carbonate and 0.5% ferric hydroxide.

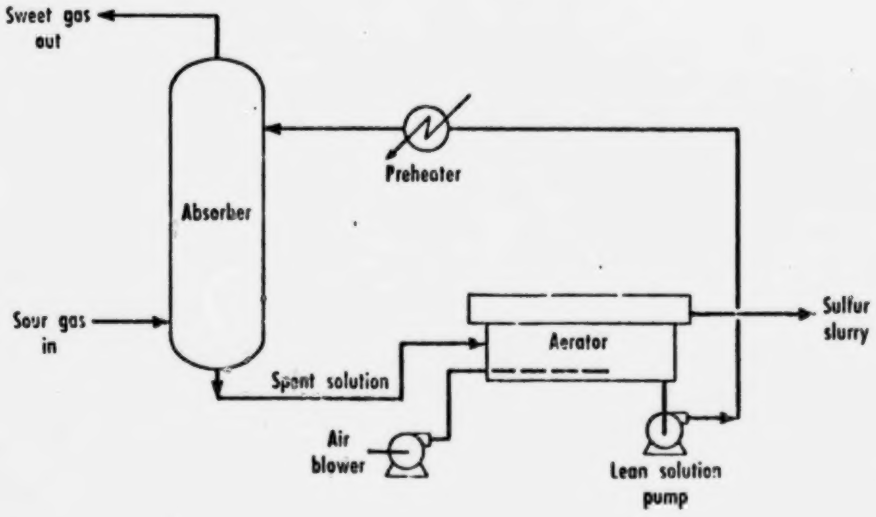


Figure 2.1.1.-43. FERROX PROCESS[49]

The solution countercurrently contacts the sour gas in the absorber. Spent solution flows from the absorber to the thionizer or regenerator, where it contacts the air and is regenerated with the formation of elemental sulfur. The sulfur is removed from the liquid surface as a froth and then passed through filters to remove the excess liquid. Regenerated solution is pumped from the thionizer through a preheater to the absorber. The filter liquid is usually discarded to purge the system of side-reaction salts.

A detailed description of equipment for and operation of the Ferrox Process is provided by Sperr[64] and by Kohl and Riesenfeld.[42]

Advantages and disadvantages of the process are given below. H_2S removal efficiency varies from 85 to almost 100%. If two-stage absorption is used, "complete" removal can be obtained. Of the H_2S removed from the gas, 70-80% can be recovered as elemental sulfur. However, there is no market for Ferrox sulfur because it is contaminated with sodium carbonate and ferric hydroxide salts. This can lead to a high chemical replacement cost for maintaining solution concentration. Since the solution is also highly corrosive, carbon steel vessels must be lined to prevent destruction.

The Ferrox Process, although mostly used for manufactured-gas applications, has been applied to selective H_2S sweetening of natural gas. New Ferrox plants have not been built because they are no longer economically competitive with other processes.

The Manchester Process differs from the Ferrox Process only in the use of multistage treating of the sour gas. Figure 2.1.1.-44 shows a schematic diagram of Manchester Process.

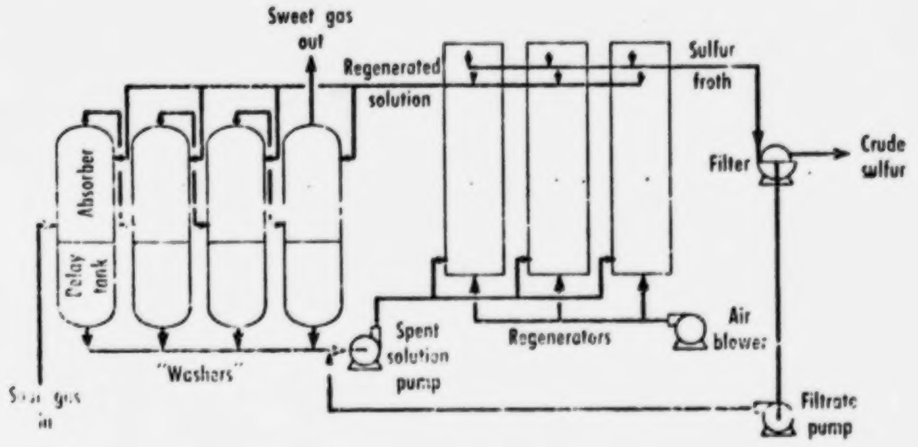

Sweet gas out
Regenerated solution
Sulfur froth
Filter
Crude sulfur
Absorber
Delay tank
"Washers"
Spent solution pump
Regenerators
Air blower
Sour gas in
Filtrate pump

Figure 2.1.1.-44. MANCHESTER PROCESS[49]

Fresh solution is fed to each "washing" stage where a single absorber is used in the Ferrox Process. Delay vessels between the absorbers and regenerators also provide holdup to ensure completion of the reaction between the alkaline sulfides and the iron oxide (Reaction 2). Regenerator towers liberate the sulfur as a froth; it is collected and filtered much as in the Ferrox Process.

Almost all Manchester treating installations have been for town-gas treating; none have been adapted for natural gas sweetening.

Although the Manchester Process can remove H_2S from gas down to undetectable quantities, the elemental sulfur is of low quality because of salt contaminants and chemical replacement costs are high. Hence, as with the Ferrox Process, more economical sweetening processes are available.

2. 1. 1. 2. 10.1.8. Lacy-Keller Process

A new, recently disclosed process[25] known as the Lacy-Keller sweetening process is now available to the industry. Details are not clear, but the process treats sour natural gas and removes both H_2S and mercaptans without removing CO_2.

Two plants employing this process have been built, one treating 12-20 million SCF/day of sour gas and recovering up to 300 lb/day of sulfur. The sour gas contains 17 grains of sulfur per 100 SCF. Initial investment cost for the plant is reportedly comparable to an amine plant designed for the same service. However, operating costs are expected to run about one-tenth of that required for the amine plant.

A sketch of the process is given in Figure 2.1.1.-45. An unidentified chemical solution contacts the sour gas in an absorber. Colloidal sulfur is produced by reaction with the sweetening agent and the solution flows to the flotation cells.

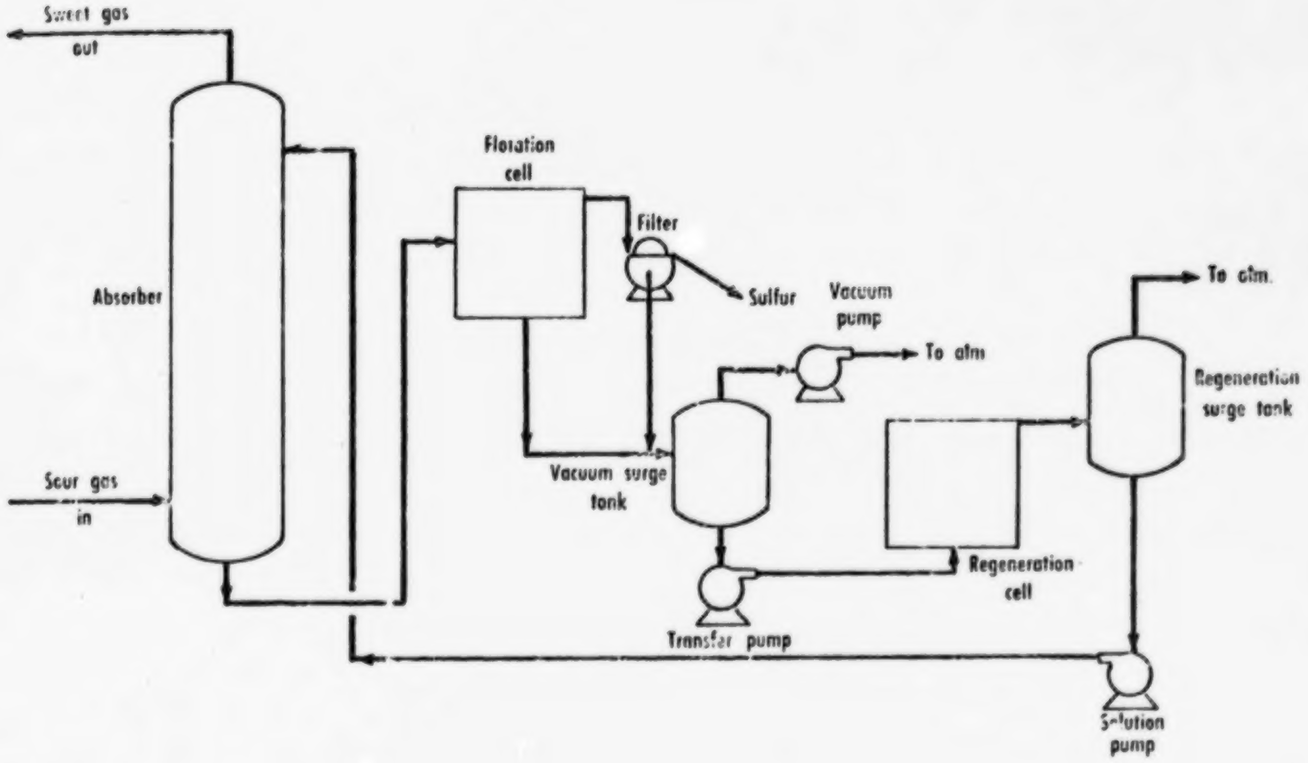

Sweet gas out
Absorber
Sour gas in
Flotation cell
Filter
Sulfur
Vacuum pump
To atm
To atm.
Vacuum surge tank
Regeneration surge tank
Regeneration cell
Transfer pump
Solution pump

Figure 2.1.1.-45. LACY-KELLER PROCESS[48]

Here, the sulfur is flocculated and carried off the surface as a thick slurry to the filter. The remaining solution is passed to a regeneration cell that operates without addition of heat and is then pumped back to the absorber. Most of the materials are of stainless steel or are lined with epoxy. Other details were not disclosed.

2.1.1.2.10.2. Investment and Operating Costs

For either technical or economic reasons, the processes included in this section are not considered useful for LNG prepurification. These reasons are summarized below:

Iron Soda:

This process was an attempt to make the iron oxide process continuous. It has found little acceptance because its operating costs are high in most applications.

Seaboard:

Developed in 1920 by the Koppers Company, this process has enjoyed widespread use where the volume of gas and the total amount of sulfur removed are large. However, the purity of the gas attainable by this process is rather low, and the economics are not too favorable when compared with the more modern processes.

Thylox:

The process is more complex than some of the others. Also, due to contamination of the elemental sulfur produced with arsenic and carbonate, the market is limited. If no credit is taken for by-product sulfur, the operating costs are high.

Fluor Solvent:

This process is in use; for pipeline gas, its economics are good. However, if the natural gas stream is rich in heavy hydrocarbons or if a high-purity treated gas is required (such as for LNG liquefaction, the process becomes less economic.

Water Wash:

This process is sometimes employed as a preliminary treating system for a very sour or acid gas, but requires a second plant (such as amine) for final cleanup.

Vacuum Carbonate:

This is a modified Seaboard Process. The gas produced is not "clean" enough for liquefaction.

Liquid Absorption-Oxidation:

The Ferrox and Manchester Processes suffer from an unmarketable product sulfur. The presence of sodium carbonate and ferric hydroxide in the sulfur eliminates its use in most industries. Also, the loss of these chemicals leads to high replacement costs.

Lacy-Keller:

This process is selective for H_2S in presence of CO_2. There are not sufficient data available for a proper evaluation of it.

2.1.1.3. Heavy Hydrocarbon Stripping

2.1.1.3.1. Molecular Sieves and Other Solid Adsorbents

Heavy hydrocarbons are recovered from natural gas streams for both economic and operational reasons. The heavy hydrocarbon components of a gas may be worth from 10 to 20 times as much when condensed and sold as a liquid as when sold as a gas. The presence of even small amounts of liquids in a pipeline can easily reduce the efficiency of gas flow by 10%, i.e., increase the pressure drop required for a given flow rate. Also, the presence of heavy hydrocarbons in the natural gas entering a liquefaction unit can result in freeze-ups in the heat exchangers or can require including additional liquid separators and special piping in the cold box to remove these materials from the process gas stream.

2.1.1.3.1.1. Process Description

Solid desiccants have a much smaller capacity for adsorbing hydrocarbons than they have for water vapor. Also, the heavier the hydrocarbon, the higher the adsorption capacity. Therefore, if heavy hydrocarbons are to be recovered by adsorption, it will be necessary to operate the adsorbent at very low mass loadings.

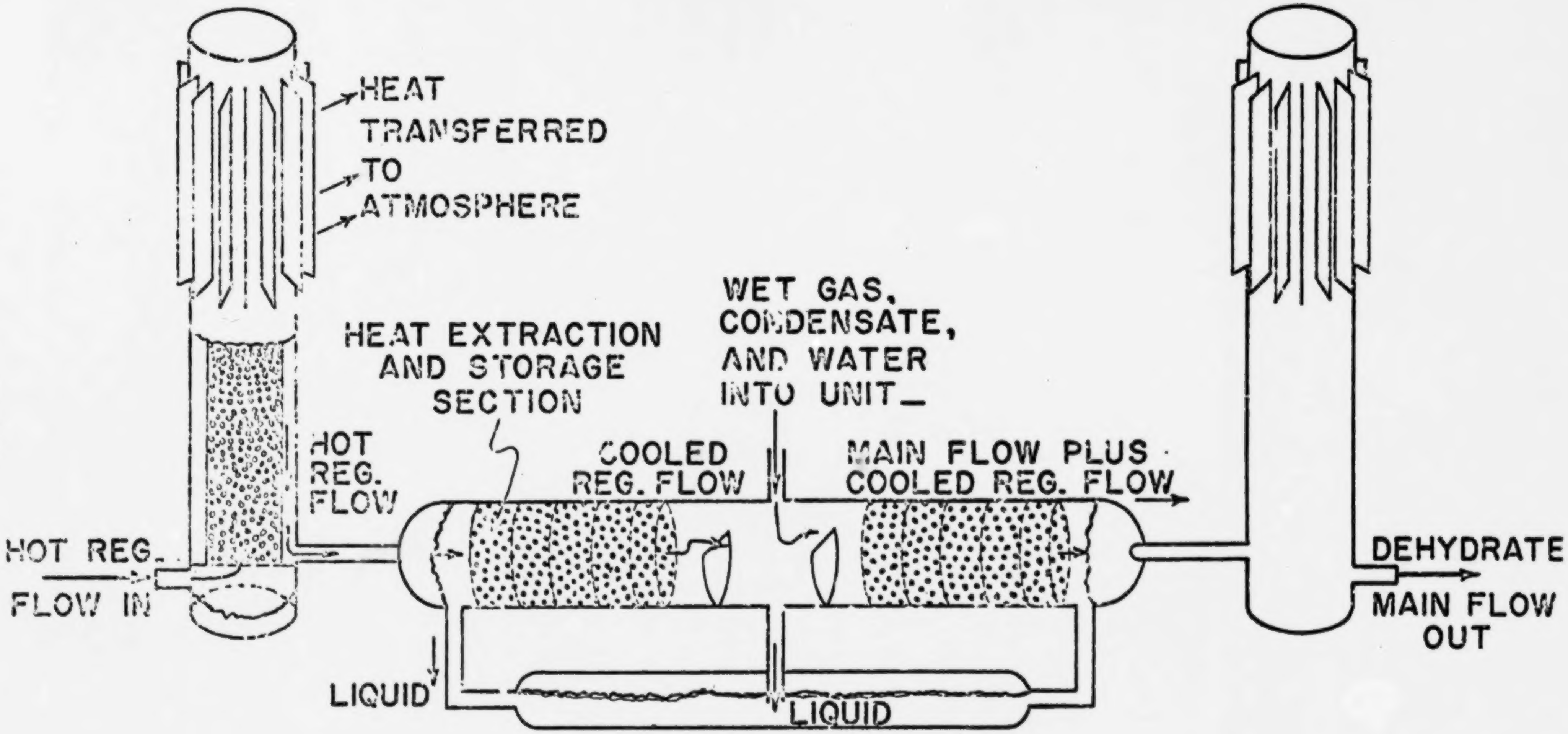

HEAT
TRANSFERRED
TO
ATMOSPHERE
HEAT EXTRACTION
AND STORAGE
SECTION
WET GAS,
CONDENSATE,
AND WATER
INTO UNIT
HOT
REG.
FLOW
COOLED
REG. FLOW
MAIN FLOW PLUS
COOLED REG. FLOW
HOT REG.
FLOW IN
DEHYDRATE
MAIN FLOW
OUT
LIQUID
LIQUID

Figure 2. 1. 1.-46. SCHEMATIC DIAGRAM OF HEAT STORAGE METHOD OF COOLING REGENERATIVE FLOW AND CONDENSING VAPORIZED WATER

The first successful and economic adsorption units for simultaneous hydrocarbon recovery and dehydration were developed by Parks and Dow.[56] A schematic diagram of the efficient regenerative unit which they developed is given in Figure 2.1.1.-46. An external heating system provides regeneration gas at 400°-450°F. Heat removed from the regeneration gas in the heat extraction and storage section is picked up by the main gas flow when the cycle is switched and flow in this unit reverses. The average cycle is about 30 minutes for adsorption and 30 minutes for regeneration and cooling. The adsorbent used is either silica or alumina gel, although some units are using molecular sieves.

2.1.1.3.1.2. Investment and Operating Costs

The investment and operating costs for the dry desiccant liquid hydrocarbon dehydration units are essentially identical with the costs of a dry desiccant dehydration unit. (See Sections 2.1.1.1.3.2. and 2.1.1.1.3.3.)

2.1.1.3.2. Conoco Adsorption Process

2.1.1.3.2.1. Process Description

The Conoco Adsorption Process is capable of economically extracting from natural gas a larger percentage of ethane and heavier hydrocarbons than is economical with most other processes in use. Investment, operating, and fuel costs are significantly lower than any other process when recovering high percentages of hydrocarbons from a relatively lean (less than 1-1/2 gallons of C_3+ per 1000 CF) gas.

The Conoco Process uses activated carbon as the adsorbent for hydrocarbon extraction. In the past this type of process has not been associated with high recoveries and has been characterized by adsorbent deterioration at an unacceptable rate. Continental's process system[13] prevents physical loss of adsorbent and loss of activity. As of this date the activated carbon beds have demonstrated a life in excess of 4 years.

The high percentage of product extraction results from the use of a closed regenerative loop that provides for a theoretical 100% recovery of the hydrocarbons desired. Practical limits of extraction have been demonstrated to be about 98-99+%, depending upon the component.

The Conoco Process is an activated carbon adsorption process when used for processing natural gas. Virtually all continuous cyclic adsorption processes consist of three repetitive cycles:

1. An adsorption cycle in which the product is removed (adsorbed) from the feed gas.
2. A regeneration cycle in which the adsorbed product is removed from the adsorbent.
3. A cooling cycle in which the adsorbent is cooled down to be ready to start Cycle 1 again.

The Conoco Process uses (as do many others) three adsorber columns. Thus, all three cycles are taking place simultaneously, each one in a different column. Operation of the process is illustrated in Figure 2.1.1.-47.

The regeneration cycle immediately follows the adsorption cycle. The Conoco regeneration cycle is conducted by circulating a hot gas through the adsorber to strip off the adsorbate. The gas leaving the adsorber is then cooled to condense the desorbed liquid hydrocarbons. Uncondensed vapors are heated and recirculated back through the adsorber.

Until now a closed regenerating cycle has not been possible because the increase in temperature plus the addition of the regenerated hydrocarbons would cause an impracticable pressure increase in the regenerating system.

In addition to the three adsorption columns, there are two VTC (volumetric thermal compensator) units. The VTC unit consists of a column of porous material. It is designed to provide the necessary compensation ad-

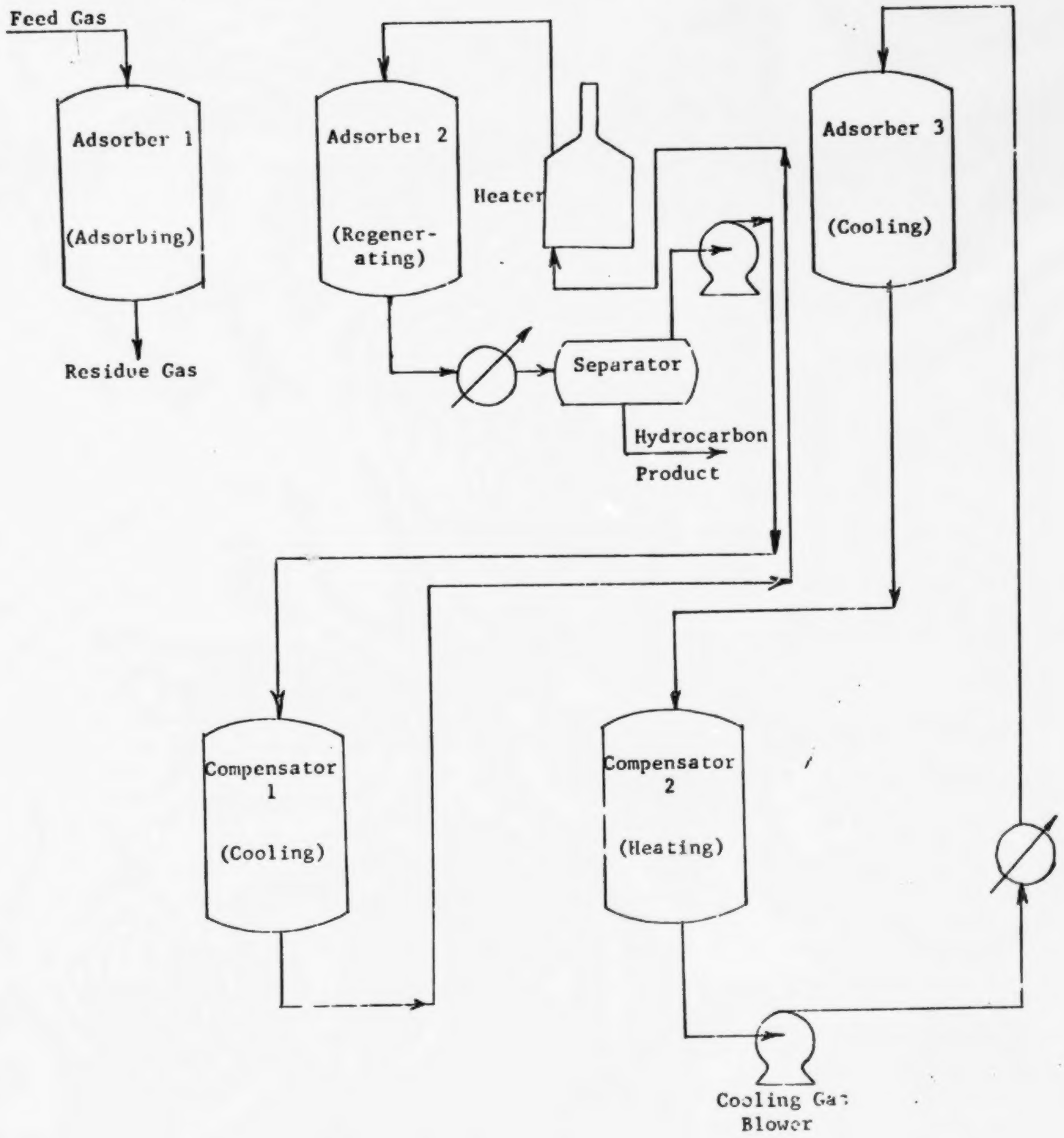

Feed Gas
Adsorber 1
(Adsorbing)
Residue Gas
Adsorber 2
(Regener-
ating)
Heater
Separator
Hydrocarbon
Product
Adsorber 3
(Cooling)
Compensator
1
(Cooling)
Compensator
2
(Heating)
Cooling Gas
Blower

Figure 2.1.1.-47. CAP IN POSITION 1 WITH COMPENSATOR AND CLOSED COOLING [13]

sorber and has the same total heat capacity as the activated carbon in each adsorber. As the adsorber is heated, the VTC unit is cooled. When properly designed, the VTC vessel compensates for the change in gas-holding capacity of the adsorber as the adsorber is heated.

Another feature of the process is that the high regeneration temperature coupled with the use of residue gas for final stripping ensures that no heavy hydrocarbons can build up on the adsorbent. This combination prevents the activated carbon from losing activity with prolonged use.

2.1.1.3.2.2. Investment and Operating Costs

We could not locate a detailed discussion of the investment and operating costs for the Conoco carbon adsorption process. The information available indicated that the costs would be similar to those for other hydrocarbon recovery systems, but cleanup of the natural gas would be more complete than obtained by the other solid adsorption or liquid absorption processes.

2.1.1.3.3. Refrigerated Separation

2.1.1.3.3.1. Process Description

This type of process should not be used for dehydration alone due to the cost. It should be used for both dehydration and recovery of natural gasoline. This process is often motivated by increased liquid recovery. Dehydration is automatically accomplished.

As shown in Figure 2.1.1.-48, all or a portion of the wellstream gas enters the coils on the bottom of the low-temperature separator, where it is used to heat the liquid and melt any hydrates formed. If the gas is not hot enough, an outside source of heat must be employed for this purpose. The partly cooled gas then flows to the heat exchanger where it is cooled to just above its hydrate point by a portion of the cold exit sales gas.

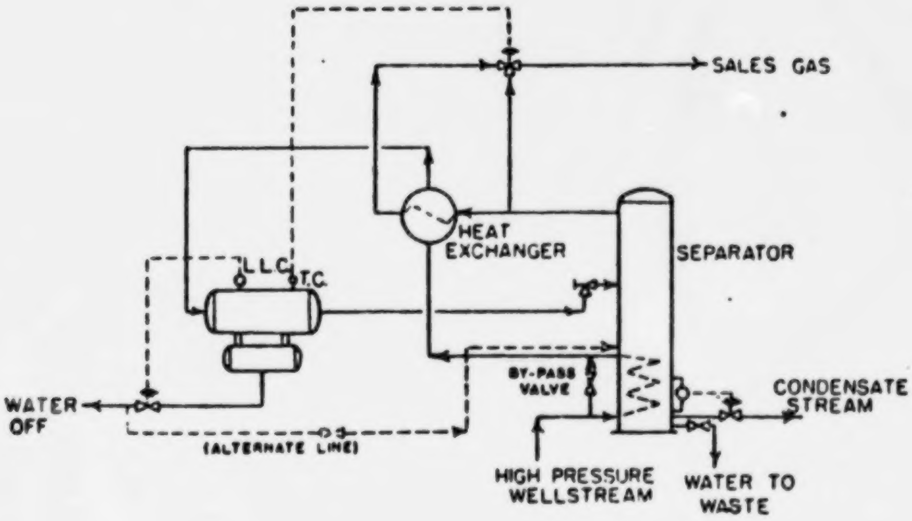

SALES GAS
HEAT EXCHANGER
SEPARATOR
L.L.C.
T.C.
BY-PASS VALVE
WATER OFF
(ALTERNATE LINE)
CONDENSATE STREAM
HIGH PRESSURE WELLSTREAM
WATER TO WASTE

Figure 2.1.1.-48. FLOW SHEET OF TYPICAL LOW-TEMPERATURE SEPARATION UNIT[11]

The liquid formed by cooling and that present in the wellstream may be handled in the knockout drum in several ways. Only the water may be removed, the condensate going across the choke with the gas as shown. An alternative method is to remove all liquids at that point and expand them separately into the bottom of the low-temperature separator.

After leaving the knockout drum, the wellstream gas is expanded into the low-temperature separator to approximately line pressure. The expansion cooling results in the condensation of a large portion of the liquefiable hydrocarbons and the formation of hydrates. After the hydrates are melted, the water and condensate are separated and either stage separation or stabilization is used to produce stock-tank liquid.

A portion of the cold sales gas from the top of the separator is used in the regeneration heat exchanger before it enters the sales line. The amount is normally controlled with a 3-way motor valve actuated by a temperature controller in the free-water knockout drum.

The only way that the liquid recovery may be accurately estimated is by analysis of the gas stream. Rather precise recovery calculations may then be made to determine the operating revenue.

In the absence of an analysis, the increased recovery may be conservatively estimated by assuming that the stock-tank recovery will be increased 0.5 bbl/10°F decrease in the separator temperature over a normal stage separator. For example, if stage separation at 80°F produces 10 bbl of condensate per million CF, low-temperature separation at 20°F will produce an additional 3.0 bbl/million CF. A further increase of not less than 10% may be expected if the low-temperature liquid is then stabilized. If recovery by stage separation is over 50 bbl/million CF, the increased recovery per 10°F decrease is usually 0.75 bbl/million CF. These figures must be recognized as pure "rules of thumb," but they do give reasonable results in many situations that have been examined.

In general, the use of low temperatures will increase the stock-tank recovery from 15 to 30% over a single stage separation. If the low-temperature fluid is stabilized, additional recoveries of 5 to 20% are further realized. After investigating the economics on numerous occasions, it appears that stabilization will give an attractive return wherever low-temperature separation is found justifiable.

The liquid recovery is a function of separation pressure as well as of temperature. The former is fixed by sales line conditions and, unfortunately, is usually above the optimum pressure for separation.

The use of a hydrate inhibitor with the expansion refrigeration minimizes 1) the hydrate temperature-control problems when cooling occurs before the choke and 2) the necessity of melting the hydrate formed.

Without an inhibitor, the temperature at the coldest point in the system is fixed by the expansion temperature drop and this hydrate point. The necessity for providing means of melting hydrates not only requires an expensive type of construction, but also brings up other process problems.

In view of these facts, and as a result of several years' observations, it has become obvious that the use of hydrate inhibitors with low-temperature units has certain advantages.

Once this fact was established, it became necessary to determine which inhibitor offered the most promise in this service. The choice was at least narrowed by experience gained in previous pipeline practice. Although it is an effective antifreeze agent, ammonia reacts with any carbon dioxide present in natural gas to form a solid carbonate that lowers efficiency and plugs lines. Calcium chloride and other chlorides are also applicable, but with them corrosion becomes a problem. By a process of elimination the alcohols, particularly methanol, have been used as hydrate inhibitors in most pipeline and gathering systems. Of these, methanol has been most used because it is both cheap and effective as an inhibiting agent.

Wide experience with the glycols in dehydrators made it obvious that at least one member of this family also showed promise. Consequently, the logical choice narrowed to the use of methanol and glycol as the hydrate inhibitor.

The flow diagram for a glycol injection system, shown in Figure 2.1.1.-49 is very similar to that used for other low-temperature systems.

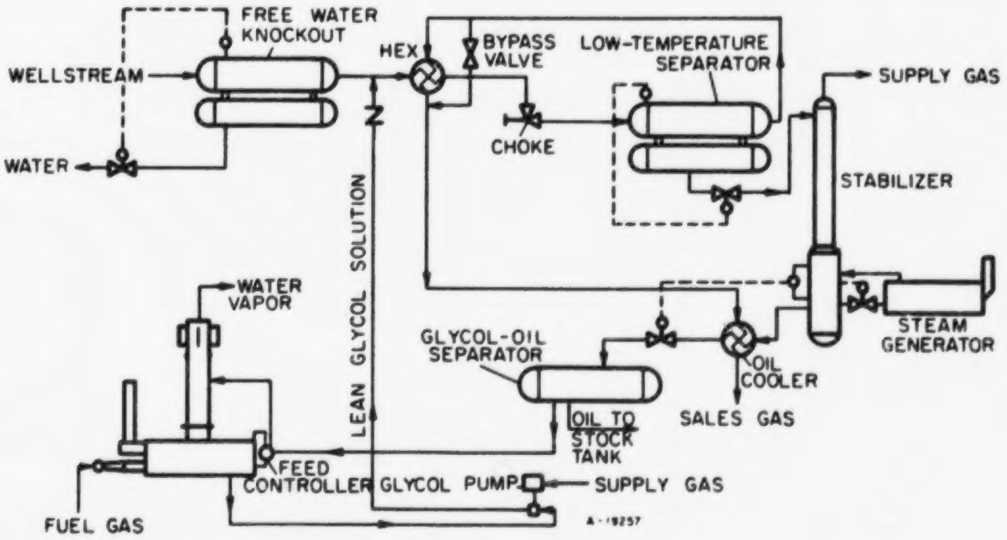

FREE WATER KNOCKOUT
WELLSTREAM
WATER
HEX
BYPASS VALVE
LOW-TEMPERATURE SEPARATOR
SUPPLY GAS
CHOKE
STABILIZER
LEAN GLYCOL SOLUTION
WATER VAPOR
GLYCOL-OIL SEPARATOR
OIL COOLER
STEAM GENERATOR
SALES GAS
OIL TO STOCK TANK
FEED CONTROLLER
GLYCOL PUMP
SUPPLY GAS
FUEL GAS
A-19257

Figure 2.1.1.-49. TYPICAL FLOW SHEET OF GLYCOL INJECTION SYSTEM WITH 125-psig WELLHEAD PRESSURE STABILIZER[10]

The inlet wellstream enters the knockout drum with or without prior cooling. Just downstream from this knockout drum but before the heat exchanger, the optimum amount of glycol is injected. This glycol then flows through the heat exchanger and across the choke into the separator with the wellstream gas. There it is separated with the condensate.

The cold sales gas then proceeds to either the wellstream heat exchanger and/or the stabilizer product cooler before it enters the sales line.

The method of separating the glycol from the condensate varies somewhat depending upon the method of handling the condensate. However, the system shown in Figure 2.1.1.-49 is typical for wellhead units with a low-pressure stabilizer.

After being separated, the glycol is reconcentrated in a unit very similar to that used for absorption dehydration and is then reinjected into the line.

A modification of the system described above has been successfully used in those cases where insufficient pressure drop is available to produce the desired refrigeration by expansion. A typical flow sheet for this system is shown in Figure 2.1.1.-50. The process factors are the same except for the source of refrigeration.

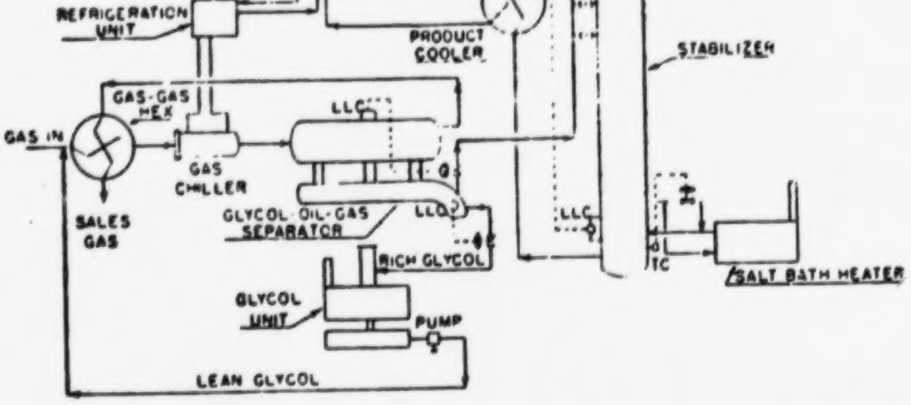

PRODUCT TO STORAGE
COOLING TOWER
GAS TC SUPPLY
REFRIGERATION UNIT
PRODUCT COOLER
STABILIZER
GAS-GAS HEX
LLC
GAS IN
GAS CHILLER
SALES GAS
GLYCOL OIL-GAS SEPARATOR
LLC
LLC
RICH GLYCOL
TC
SALT BATH HEATER
GLYCOL UNIT
PUMP
LEAN GLYCOL

Figure 2.1.1.-50. TYPICAL FLOW SHEET FOR MECHANICAL REFRIGERATION SYSTEM WITH STABILIZER[10]

The use of mechanical refrigeration is not advisable for dehydration unless the liquid recovery warrants it. The dehydration accomplished, however, plays an important role in comparing costs with an oil absorption plant where other dehydration facilities are necessary.

Simple compression and cooling have been widely used for recovery, but this is a relatively inefficient process and has been abandoned in new designs, except for small plants (Figure 2.1.1.-51) where the cost must be kept low. Adsorption on activated charcoal was practiced to some extent, but this process has become insignificant because of such troubles as the dusting away of the adsorbent and a gradual loss of ability to adsorb that is caused by the collection of heavy oil on the surface.

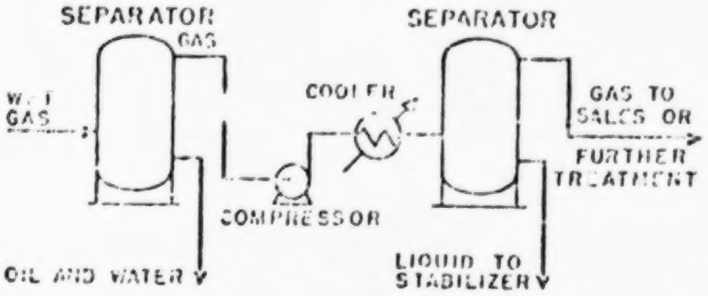


Figure 2.1.1.- 51. RECOVERY OF NATURAL GAS LIQUIDS BY COMPRESSION

Most modern plants employ the oil absorption process. This process consists of having natural gas countercurrently contact a hydrocarbon oil, such as kerosine, naphtha, or mineral seal oil. All of the hydrocarbons in natural gas dissolve in this type of oil to some extent, but the degree increases with increase in molecular weight. Dry gas leaves the top of the contactor (absorber), passes through a mist-remover, and is then dehydrated if it is to be sent to a transmission line. Rich oil from the bottom of the absorber is sent through heat exchangers to a stripper for regeneration. Light components are removed from the rich oil in the stripper by having it countercurrently contact open steam. The regenerated lean oil is removed from the stripper, cooled, and reused. The overhead product from the still is condensed and separated from water. This material is raw or "wild" gasoline, and must be "stabilized" by fractional distillation to remove components that are too volatile — i.e., the methane, ethane, propane, and all or part of the butane.

It is usual to recover mixed propane and butane from raw gasoline as LPG or to recover propane and butane separately. In fact, where there is a market for isobutane, it can be separated from normal butane by fractional distillation in a "butane splitter." Many plants make special products; in some new plants, a high-ethane stream is recovered as a raw material for petrochemical manufacture. Low-temperature fractional distillation has been employed for this purpose.

Some of the essential features of a typical adsorption-type natural gasoline plant are indicated in the flow sheets in Figures 2.1.1.-52 and 2.1.1.-53.

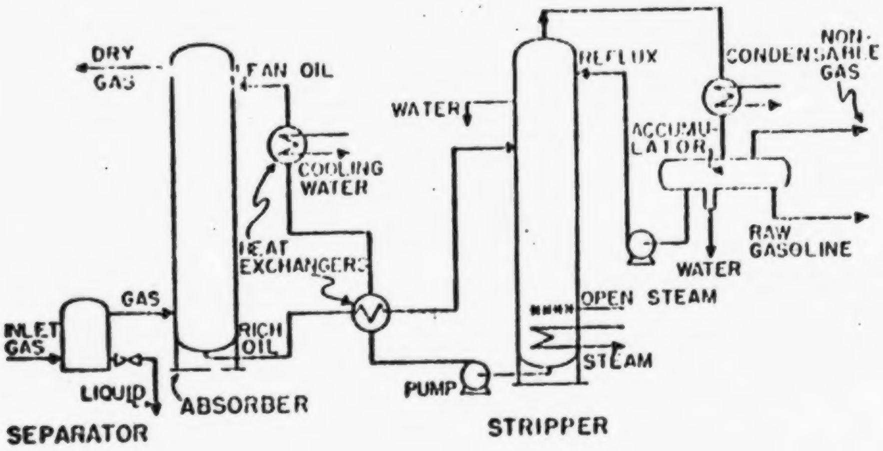

DRY GAS
LEAN OIL
COOLING WATER
HEAT EXCHANGERS
GAS
INLET GAS
RICH OIL
LIQUID
SEPARATOR
ABSORBER
WATER
REFLUX
CONDENSER
NON-CONDENSABLE GAS
ACCUMULATOR
RAW GASOLINE
WATER
OPEN STEAM
STEAM
PUMP
STRIPPER

Figure 2.1.1.- 52. SECTION OF NATURAL GASOLINE PLANT

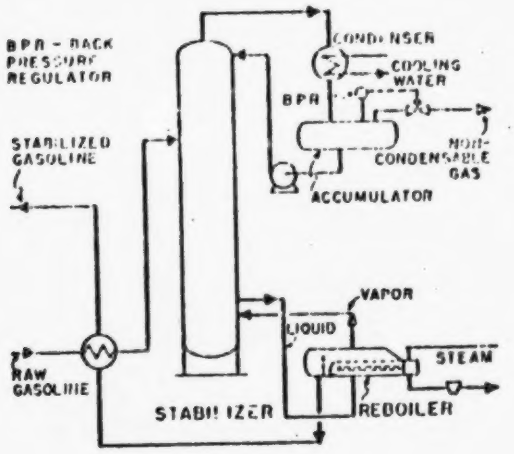

BPR - BACK PRESSURE REGULATOR
CONDENSER
COOLING WATER
BPR
STABILIZED GASOLINE
NON-CONDENSABLE GAS
ACCUMULATOR
VAPOR
LIQUID
STEAM
RAW GASOLINE
STABILIZER
REBOILER

Figure 2.1.1.-53. STABILIZATION OF RAW GASOLINE

Many modern gasoline plants operate with absorber pressures as high as 500-800 psi, although earlier units were designed for considerably lower pressures, even as low as 30 psi. Cycling plants that treat high-pressure gas frequently operate with absorber pressures of 1200-1800 psi, and occasionally more; however, troubles, such as loss of some absorber oil by retrograde vaporization, may be encountered at pressures exceeding 800 psi.

Where the field pressure is less than 30-50 psi, the gas is cleaned and compressed to at least this pressure, and preferably more, before processing. If the gas is to be sent to a high-pressure transmission line, absorption at substantially line pressure is indicated. Gas from deep wells is under high pressure initially and may be reduced in pressure at the wellhead or, at least, before being sent to the inlet separator at the plant.

Plants were originally designed for recoveries of as low as 40% of the butanes and of from 90 to 95% of the pentanes. Later plants have been designed for recoveries of 40% or more propane, 80-90% butanes, and 99% C_5+. These figures cannot be specified independently; for a given gas composition, fixing the operating conditions for a given recovery of one component fixes the recoveries of the other components. The degree of absorption is greater the higher the pressure and the lower the temperature. A common lean oil temperature at the inlet to the absorbers in the Southwest during the summer is 90°F, which is about the lowest practically obtainable with the cooling water available. Water can be cooled to within 5°-10°F of the wet-bulb temperature of the air, and oil can be cooled to within 5°-10°F of the cooling water.

Chilling of lean oil to 60°F by propane refrigeration is currently gaining in popularity because of decreased oil handling and more favorable relative solubilities of the light components; relatively more of the desirable components dissolve at lower temperatures.

Selection of absorber temperature and pressure, the ratio of lean oil to gas to be employed, and the number of contacting trays to be used are fixed by chemical, engineering, and economic considerations. Absorbers commonly have 24-36 plates that are spaced 18-24 in. apart. Some important factors affecting absorber design are the gas pressure at the plant inlet, the composition of the gas to be treated, the final pressure required for the treated gas, and the lowest temperature available for lean oil.

The higher the absorber pressure employed, the smaller is the required vessel, but thicker walls must be used. Higher pressure means higher liquid pumping costs. Too high a pressure may lead to unfavorable effects, such as retrograde vaporization of absorption oil and solution of too much of the light components. Expensive refrigeration is required for chilling lean oil to temperatures less than 80°-90°F. Increased lean oil rate means greater cost for regeneration, cooling, and recirculation, and an increased number of plates means a higher initial equipment cost. In general it is cheaper to buy a taller column than to pay for increased heating and cooling of lean oil. All of these factors must be taken into consideration to determine the optimum design and operating conditions — the conditions that lead to the desired result at the least expense.

When absorber pressure is high, the rich oil may be flashed in one or more stages of about 50% pressure reduction per stage before it is sent to the stripper for regeneration. Since strippers act in the opposite sense to the absorbers, they function better the lower the pressures and the higher the temperature. Stripper pressure cannot be too low, however, as some valuable material will not be condensed, and so will be lost in the overhead product. The available condenser temperature, which is determined by the temperature of the cooling medium (normally water), fixes the pressure required for condensation. In accordance with the principles explained for vapor pressure, if the temperature is fixed, the dew and bubble point pressures are fixed. The higher the temperature, the higher these pressure values are. The bubble point pressure must be reached for complete condensation. Many units operate at a pressure between the dew and bubble point pressures, with the result that condensation is not complete and uncondensed vapor, composed largely of lighter components, is vented from the condenser to the fuel-gas-gathering system.

Flashing makes possible the treatment of both high- and low-pressure gas in one plant, such as illustrated in the flow sheet in Figure 2.1.1.-54.

Flashing not only reduces the pressure on the rich oil to be stripped, but, as for field separators, the reduction in pressure releases from solution lighter components and thereby lightens the load on the still and other equipment.

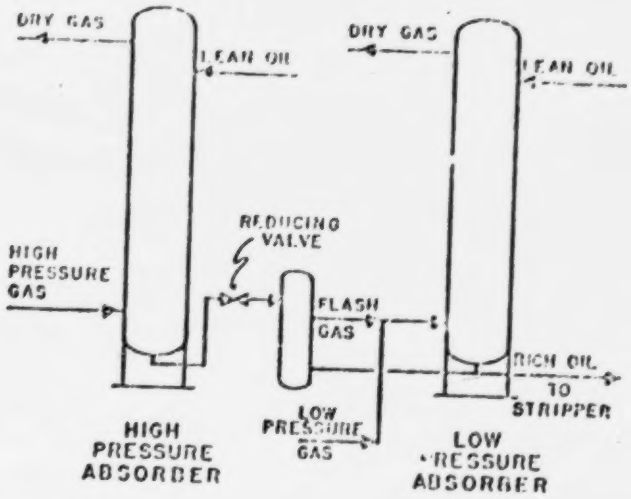


Figure 2.1.1.-54. FLASHING OF RICH OIL

Flash gas is largely methane and ethane, but contains some of the heavier components (propane and butane). Before this gas is sent to the pipeline or used as plant fuel (some gas is invariably consumed for making steam, running pumps, or heating certain streams of materials), it may be treated in a separate absorber to reduce its propane and butane content. Such an absorber treating only flash gas is called a reabsorber. In some installations flash gas is compressed and sent back to the original absorber; here the compressor is called a recompressor.

In the stripper, the LPG and gasoline components dissolved in the absorption oil are removed. Open or "agitation" steam is blown upward through a plate column, contacting the descending rich oil. Steam serves as a hot inert gas; the components of the rich oil tend to vaporize into it. The absorption oil employed should not be too volatie or it will vaporize appreciably. Absorber oil is usually a mineral seal oil with a molecular weight of 180-220. Losses as low as 0.1 gal/million CF of gas treated have been claimed. The hydrocarbons dissolved in the absorption oil vaporize into the steam, causing regeneration of the lean oil by "stripping" or "denuding" the rich oil of these components. The lean oil is removed from the bottom of the tower, cooled, and sent to a surge tank, from which it is recirculated to the absorber(s).

In this process all components of the rich oil tend to evaporate into the rising steam stream until their partial pressures in the vapor correspond to their vapor pressures at the operating temperature. This tendency cannot be fully achieved in practice, but can be more readily approached the greater the number of plates. Some steam is condensed in the stripping column, particularly in the upper part, from which it is removed from wells in one or more plates. Condensation of steam is completed in the condenser, and the water, which is immiscible with raw gasoline, settles into and is removed from a leg in the accumulator just beyond the condenser. Condensation of the stripping agent is very important, as it increases the partial pressures of the remaining components of the overhead vapor and thus permits more complete condensation of the hydrocarbons.

Strippers are commonly operated at pressures as high as 75-100 psi. Lower pressure favors greater stripping action, but makes condensation in the overhead condenser more difficult. Heat is required for vaporization in the stripping column. Some heat comes from the open steam. To conserve on the amounts of open steam required, it is usual to preheat the rich oil entering the column and to heat the bottoms by closed steam coils or direct firing by gas so that the bottom temperature is 400°-500°F.

Most stripping stills are provided with reflux to fractionally distill any vaporized absorber oil out of the overhead stream. This practice requires the use of a reboiler – a heater at the bottom of the column – since the heat load on the column is increased. When reflux is used, the still serves two purposes and is commonly called an "evaporator." Dual stills (Figure 2.1.1.- 55) are commonly used in high-pressure cycling plants, with the primary unit operating at 125-225 psi, and the secondary at atmospheric to 75 psi. The pressure in the primary unit is too high to permit complete stripping. Enough steam can be used in the high-pressure still to strip down to about butane and so fix the end point of the gasoline recovered in the next still. The overhead from the primary still is very volatile because of its methane, ethane, and LPG content, and cannot be completely condensed at

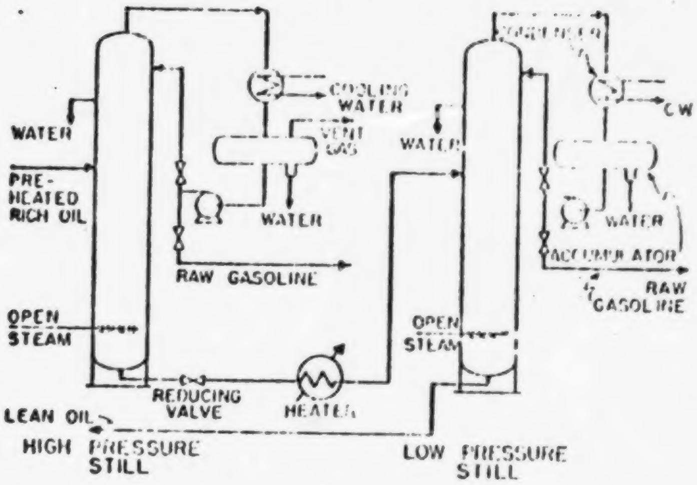

WATER
PRE-HEATED RICH OIL
OPEN STEAM
COOLING WATER
VENT GAS
WATER
RAW GASOLINE
REDUCING VALVE
HEATER
LEAN OIL
HIGH PRESSURE STILL
CONDENSER
CW
WATER
WATER
ACCUMULATOR
RAW GASOLINE
OPEN STEAM
LOW PRESSURE STILL

Figure 2.1.1.-55. DUAL STILL FOR LEAN OIL REGENERATION

moderate pressure. Finished gasoline can be removed from the partially regenerated absorption oil in the low-pressure still. Since the lighter components will have been removed in the primary still, the secondary still overhead can be completely condensed at moderate pressure.

2. 1. 1. 3. 3. 2. Investment Costs

The initial cost may be estimated from Figure 2.1.1.-56, although the figures shown do not include any equipment for processing the liquid recovered. The prices shown may be used for the preparation of preliminary cost estimates in the comparison of the various systems. Separators with a 1050-psig working pressure were chosen for study because they are most common. For intermediate knockout drum pressures the curves may be interpolated.

The initial cost of the type of unit shown in Figure 2.1.1.-49, complete with a low-pressure stabilizer, may be estimated from Figure 2.1.1.-57. The prices shown are for a 1000-psig sales-line pressure. For a sales-line pressure of 500 psig or less, about $500 may be subtracted from the total unit cost while the same figure should be added for 1500-psig working pressure.

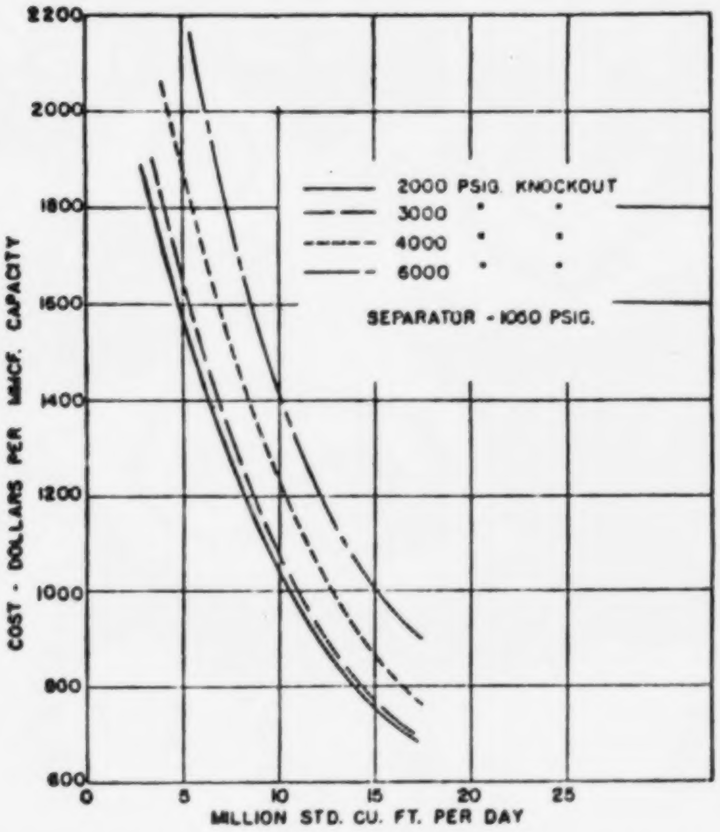


Figure 2.1.1.-56. APPROXIMATELY INITIAL EQUIPMENT COST OF LOW-TEMPERATURE SEPARATION EQUIPMENT [11]

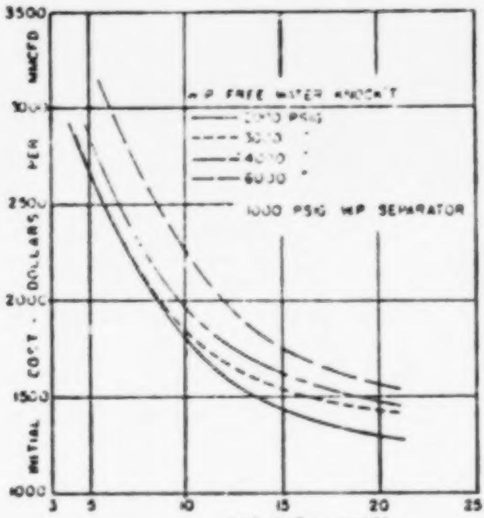


Figure 2.1.1.-57. INITIAL COST ESTIMATION CURVES FOR GLYCOL INJECTION SYSTEMS, INCLUDING STABILIZER[10]

The approximate prices are for what might be considered standard units for the conditions shown. For unusual applications, very rich gas, or where extra heat exchange is needed, the prices shown will be slightly low. However, they may be used for estimation purposes in most instances.

For larger central plants up to 100 million CF/day capacity other than those shown, a price of $1400/million CF is satisfactory. Above 100 million CF/day capacity, a price of $1250/million CF/day should be adequate. The prices shown are for equipment only since installation costs will vary with the location.

2.1.1.3.3.3. Operating Costs

The operating cost of low-temperature equipment varies from $0.20 to $9.60/million CF/day in most instances, including the liquid-handling equipment. These costs apply to leases where no unusual operating problems have existed and where only daily inspection has been practiced.

2.1.1.3.4. Parkersburg Hyreco Cold Separation Unit

2.1.1.3.4.1. Process Description

The Parkersburg Hyreco Cold Processing Unit is a complete and unitized high-pressure gas well hookup. It consists of 1) a high-recovery separator, 2) a heat exchanger, 3) a wellhead knockout drum, 4) controls, 5) a second-stage flash chamber (when required), and 6) prefabricated piping.

The wellhead processing unit flow pattern is presented in Figure 2.1.1.-58.

The gas and oil fluid from the wellhead(1) pass into the high-recovery separator (2) through a coil in the warm zone of the unit. By the time the wellstream reaches the coil in the warm zone of the separator, some of the wellstream gas has condensed into a liquid and a great amount of the total water saturating the gas is liquefied. As the wellstream passes through, the heat from the wellstream performs an important warming job by preventing the hydrocarbons and water inside the high-recovery separator from freezing. The wellstream passes into the heat exchanger (3), a shell and tube arrangement in which the warm wellstream passes through small internal tubes where the temperature (4) is dropped to a safe operating point above the hydrate point. It then passes into a vertical liquid knockout drum (5), a separating device which removes all liquid water and hydrocarbon fluids that are present in the gas at this pressure and temperature.

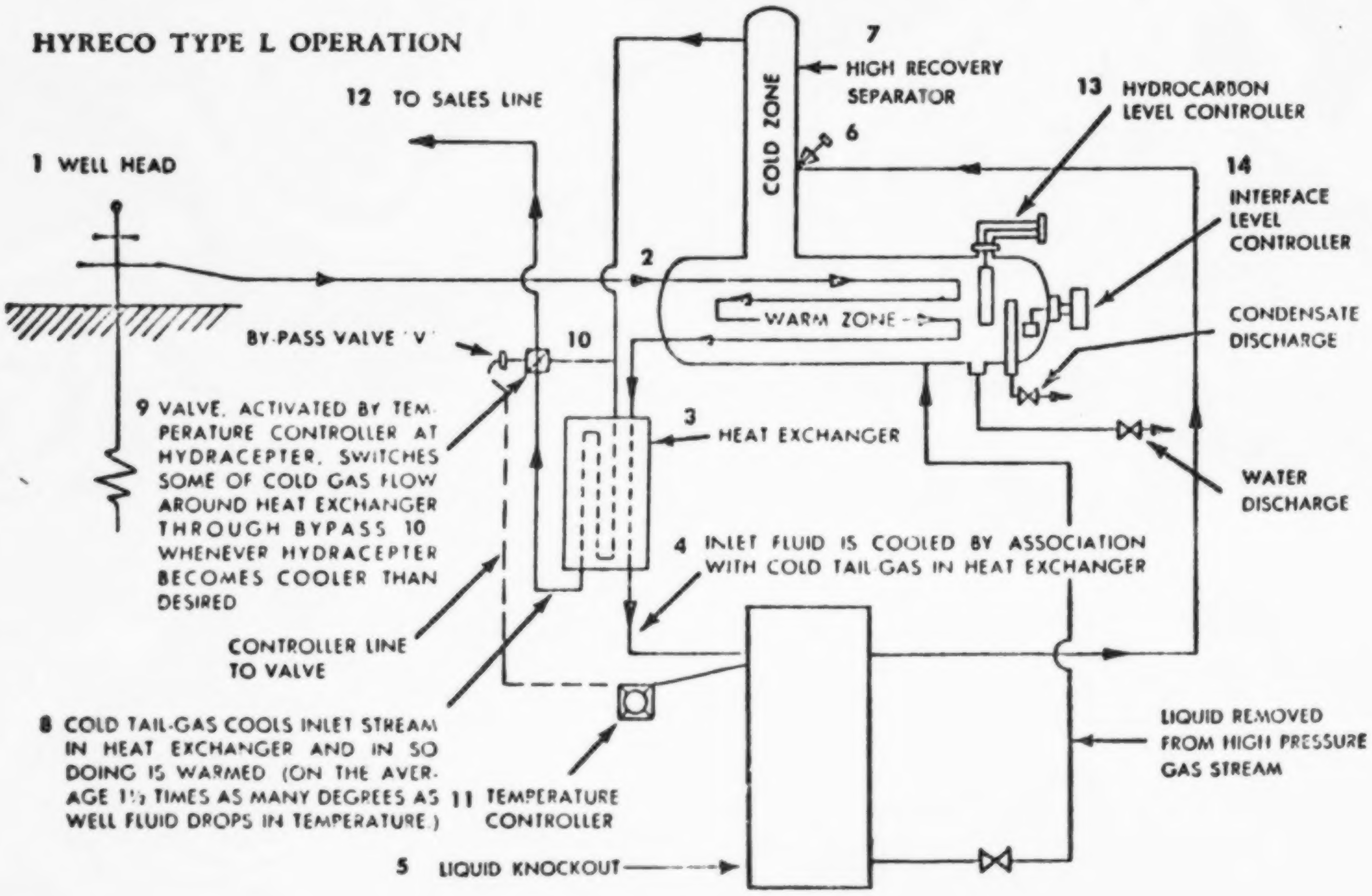

HYRECO TYPE L OPERATION
12 TO SALES LINE
1 WELL HEAD
7
HIGH RECOVERY SEPARATOR
COLD ZONE
6
13 HYDROCARBON LEVEL CONTROLLER
14 INTERFACE LEVEL CONTROLLER
2
WARM ZONE
CONDENSATE DISCHARGE
BY-PASS VALVE 'V'
10
9 VALVE, ACTIVATED BY TEMPERATURE CONTROLLER AT HYDRACEPTER, SWITCHES SOME OF COLD GAS FLOW AROUND HEAT EXCHANGER THROUGH BYPASS 10 WHENEVER HYDRACEPTER BECOMES COOLER THAN DESIRED
3 HEAT EXCHANGER
WATER DISCHARGE
4 INLET FLUID IS COOLED BY ASSOCIATION WITH COLD TAIL-GAS IN HEAT EXCHANGER
CONTROLLER LINE TO VALVE
8 COLD TAIL-GAS COOLS INLET STREAM IN HEAT EXCHANGER AND IN SO DOING IS WARMED (ON THE AVERAGE 1½ TIMES AS MANY DEGREES AS WELL FLUID DROPS IN TEMPERATURE.)
11 TEMPERATURE CONTROLLER
LIQUID REMOVED FROM HIGH PRESSURE GAS STREAM
5 LIQUID KNOCKOUT

Figure 2.1.1.-58. FLOW PATTERN OF WELLHEAD PROCESSING UNIT

As the liquid-free gas stream passes into the cold zone (7) of the high-recovery separator, the gas is expanded through a choke (6). This expansion causes a drop in pressure with a corresponding drop in temperature, which condenses the remaining vapor hydrocarbons. Most of the vapor water left in the system is also condensed by this sudden drop in pressure and temperature. The wellstream then passes into the high-recovery separator tower, which produces dry gas, liquid hydrocarbons, and water. These liquids drop to the warm zone of the high-recovery separator, where the warm coil does its important heating job of preventing hydrate formation. By maintaining a warm liquid in this zone, the hydrates are melted and rendered harmless to subsequent operations. At this point, the condensates are automatically tapped off to the tanks or to a second-stage separator. The cold, dry sales-line gas is routed back through the shell side of the heat exchanger (8) and is warmed by the incoming wellstream.

In turn, the cold gas cools the incoming wellstream. A bypass valve (9) is activated by a temperature controller (11) connected to the vertical knockout drum. The temperature of the incoming stream at the knockout drum must be maintained at a safe margin above the hydrate point. If the gas temperature is lower than desired, the control diverts a portion of the cold gas flow from the heat exchanger through the bypass (10) to help maintain a safe temperature. The gas is then passed through piping (12) to the sales line.

2.1.1.3.4.2. Investment Costs

The initial investment cost may be estimated from the table given below,[11] which have been adjusted to 1968 costs.[54]

	Inlet Pressure, psig			
Capacity, 10^6 CF/day	2000	3000	4000	6000
	Cost, $/$10^6$ CF/day			
5	3400	3500	4000	4600
10	2200	2300	2600	3000
15	1600	1650	1900	2100

These costs do not include any equipment for processing the recovered liquid because both the type and number of stages utilized in stabilization will depend to a large extent on the market available. The low-temperature separator has a 1050-psig working pressure, which is common throughout the industry. The knockout drum design pressure is the heading for the price column, with price increasing with increasing pressure.

2.1.1.3.4.3. Operating Costs

The operating cost of low-temperature equipment operated to recover liquid hydrocarbons varies from $0.20 to $0.60/million CF including the liquid handling equipment.[11] These costs apply to leases where unusual operating problems are not encountered.

Heavy hydrocarbon recovery is generally increased by 15-30% by the use of low-temperature separation instead of single-stage flash. This recovery can be further enhanced by an additional 5-20% by stabilizing the recovered liquid instead of flashing to stock tank conditions. The complete economics need to be examined for each particular case, but it generally appears that low-temperature separation, where applicable, plus stabilization of the recovered liquid will give an attractive return on investment.

2. 1. 1. 4. Nitrogen Removal

2. 1. 1. 4. 1. Process Description

2. 1. 1. 4. 1. 1. Recirculating Nitrogen Cycle

A schematic diagram of a process patented by J. L. Cost[14] for the removal of nitrogen from natural gas is presented in Figure 2.1.1.-59. In this design the required refrigeration load is supplied by isentropic expansion of nitrogen vapor through a series of expansion turbines. Part of this high-pressure nitrogen comes from the methane-nitrogen feed mixture as the overhead product of the distillation column, and the remainder from a recirculated nitrogen stream. The separated nitrogen is thus used to produce a portion of the refrigeration required.

When the natural gas supply pressure is less than the system operating pressure, the natural gas is compressed by the feed compressor and aftercooled in Heat Exchanger A. If the feed pressure is sufficiently above system operating pressure, the natural gas is expanded through the feed expansion turbine to the required pressure, and then passes through Precooler B, where it is cooled by the separated methane. It next passes to the product vaporizer, where it is partially liquefied and furnishes the heat of vaporization of the methane product, and then to the reboiler for the distillation column, where it provides the heat required for column operation and, in turn, is completely liquefied. The liquefied feed is subcooled by countercurrent heat exchange with cold nitrogen vapor in Heat Exchanger C and is then throttled through a valve to the column operating pressure and injected near the middle of the distillation column. Vaporized nitrogen is taken off as column overhead, and liquid methane as bottoms product. Gaseous methane from the product vaporizer passes through Feed Precooler B to the methane product compressor and aftercooler. The gaseous nitrogen overhead passes through a partial condenser (reflux condenser). The resulting liquid nitrogen is separated from the nitrogen vapor and returned to the top of the column.

Uncondensed nitrogen from the separator joins the recirculated nitrogen stream at E. From E the combined stream passes to Liquid Feed Subcooler C, and then through Expansion Turbine G, where the temperature and pressure of the stream are reduced. The cooled nitrogen passes through the coils in the partial condenser, a separate coil in Liquid Feed Subcooler C, a

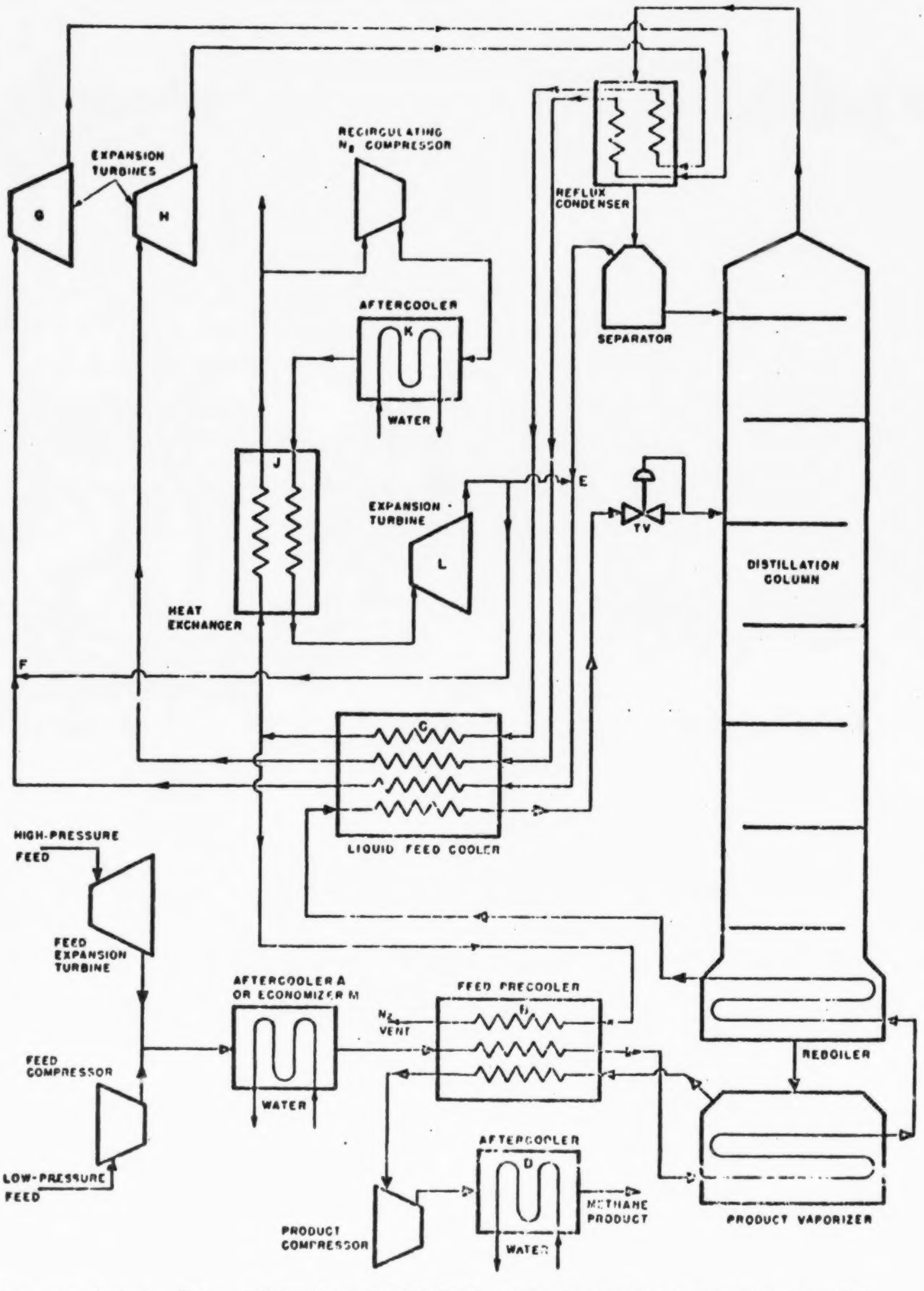

EXPANSION TURBINES
G
H
RECIRCULATING N_2 COMPRESSOR
AFTERCOOLER
K
WATER
REFLUX CONDENSER
SEPARATOR
J
HEAT EXCHANGER
EXPANSION TURBINE
L
E
TV
DISTILLATION COLUMN
F
C
LIQUID FEED COOLER
HIGH-PRESSURE FEED
FEED EXPANSION TURBINE
FEED COMPRESSOR
LOW-PRESSURE FEED
AFTERCOOLER A OR ECONOMIZER M
WATER
N_2 VENT
FEED PRECOOLER
B
REBOILER
PRODUCT VAPORIZER
AFTERCOOLER
D
METHANE PRODUCT
PRODUCT COMPRESSOR
WATER

Figure 2.1.1.-59. COST'S RECIRCULATING NITROGEN PROCESS FOR METHANE-NITROGEN SEPARATION[14]

second expansion turbine (H), and back through a second coil in the partial condenser. From the reflux condenser it again passes through Liquid Feed Subcooler C. The stream is divided and part goes through Feed Precooler B, where it supplies a portion of the cooling required by the incoming feed gas, and is then vented to the atmosphere. The remainder of the stream passes through Heat Exchanger J, where it cools the compressed recirculated nitrogen stream while being warmed to about 70°F; then part of it is vented to the atmosphere, while that required for recirculation is sent to the recirculating nitrogen compressor.

The compressed recirculating nitrogen stream is cooled by water in Aftercooler K and then by the low-pressure nitrogen stream in Exchanger J. The cooled compressed nitrogen is expanded through Turbine L to the pressure and temperature of the nitrogen vapor leaving the partial condenser, and joins with the latter vapor stream at E. Under certain operating conditions it is possible to have the recirculated nitrogen stream join the nitrogen vapor stream after the first pass through Liquid Feed Subcooler C, at point E.

2.1.1.4.2. Ammonia-Ethylene-Methane-Cascade Cycle

Two processes for separating nitrogen from natural gas, utilizing the latent heat of vaporization of liquid methane to provide the refrigeration requirements, are presented schematically in Figures 2.1.1.-60 and 2.1.1.-61. The refrigeration available from the separated nitrogen is used to supply a portion of the reflux condenser requirements and to remove heat from the feed gas.

The pressure at which the supply of natural gas is available to the plant and the condition of the feed entering the column influence the choice of flow paths through the system and the operation of the methane refrigeration

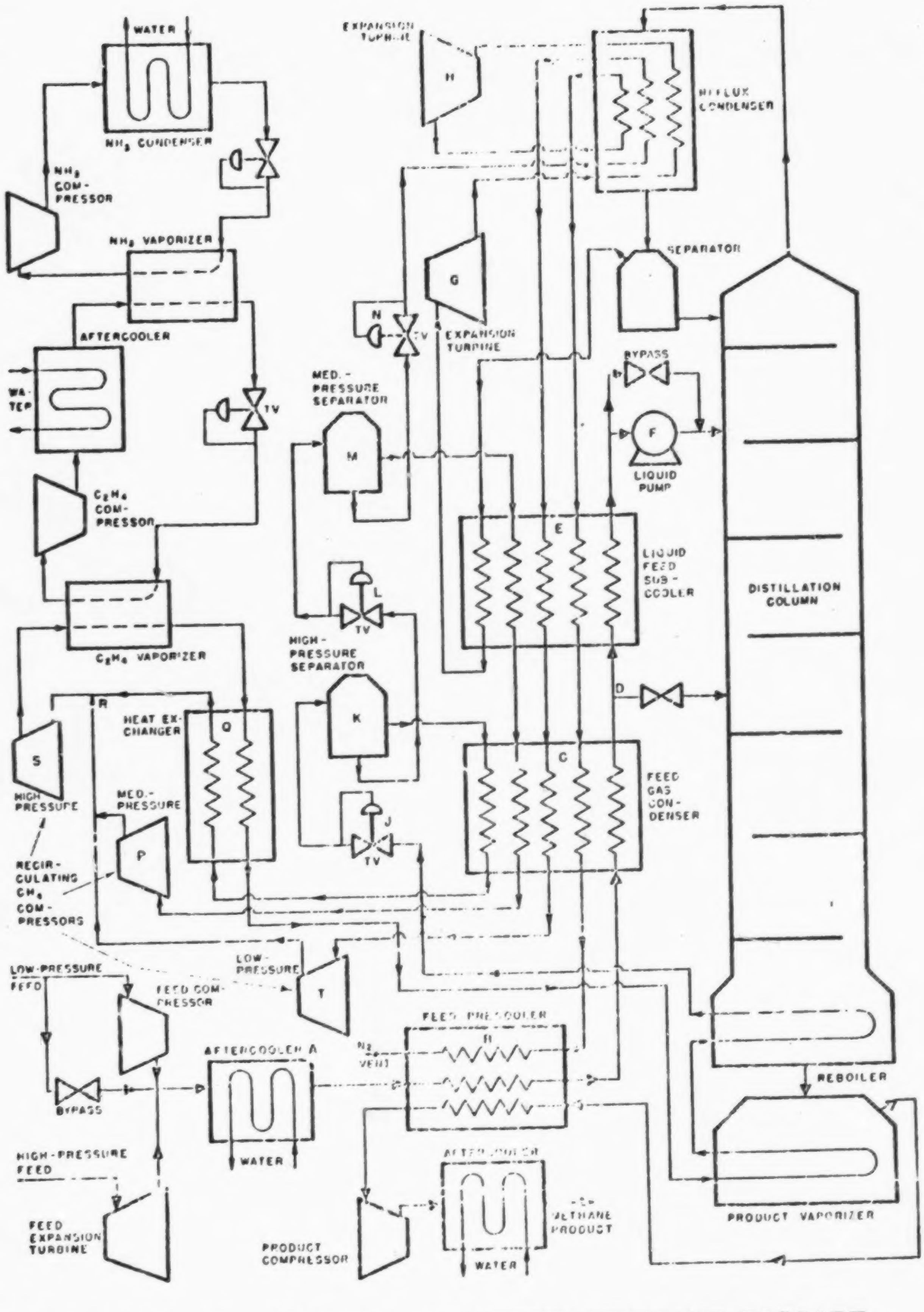

WATER
NH₃ CONDENSER
NH₃ COMPRESSOR
NH₃ VAPORIZER
AFTERCOOLER
WATER
TV
C₂H₄ COMPRESSOR
C₂H₄ VAPORIZER
R
HEAT EXCHANGER
Q
S
HIGH PRESSURE
MED.-PRESSURE
P
RECIRCULATING CH₄ COMPRESSORS
LOW-PRESSURE
T
LOW-PRESSURE FEED
FEED COMPRESSOR
BYPASS
AFTERCOOLER A
WATER
HIGH-PRESSURE FEED
FEED EXPANSION TURBINE
EXPANSION TURBINE
H
REFLUX CONDENSER
G
N
TV
EXPANSION TURBINE
SEPARATOR
MED.-PRESSURE SEPARATOR
M
BYPASS
F
LIQUID PUMP
E
LIQUID FEED SUB-COOLER
DISTILLATION COLUMN
L
TV
HIGH-PRESSURE SEPARATOR
K
D
C
FEED GAS CONDENSER
J
TV
FEED PRECOOLER
B
N₂ VENT
REBOILER
PRODUCT VAPORIZER
AFTERCOOLER
METHANE PRODUCT
PRODUCT COMPRESSOR
WATER

Figure 2.1.1.-60. AMMONIA-ETHYLENE-METHANE CASCADE PROCESS FOR METHANE-NITROGEN SEPARATION

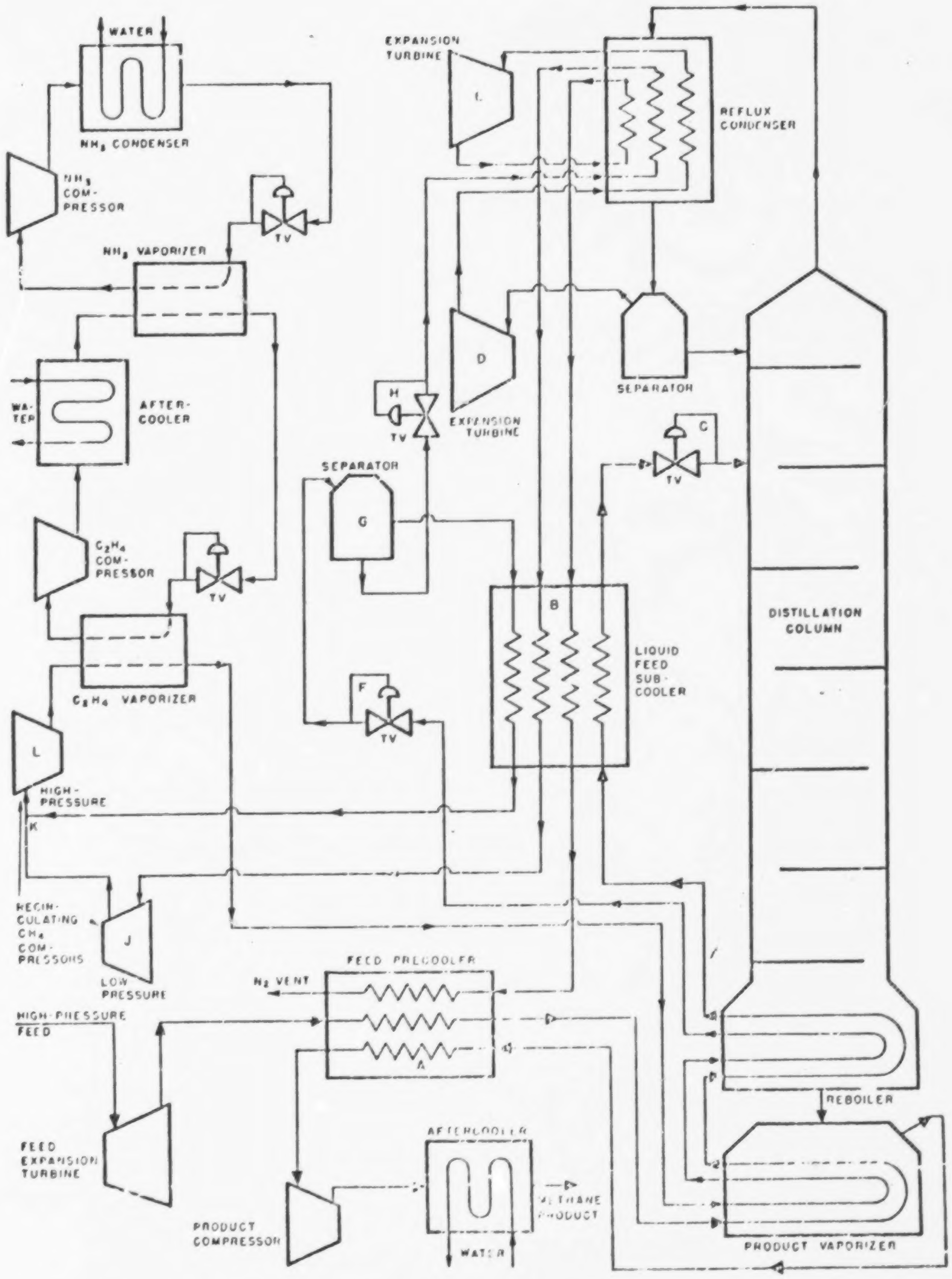

WATER
NH_3 CONDENSER
NH_3 COM-PRESSOR
TV
NH_3 VAPORIZER
WA-TER
AFTER-COOLER
C_2H_4 COM-PRESSOR
TV
C_2H_4 VAPORIZER
L
HIGH-PRESSURE
K
RECIN-CULATING CH_4 COM-PRESSORS
J
LOW PRESSURE
HIGH-PRESSURE FEED
FEED EXPANSION TURBINE
EXPANSION TURBINE
E
REFLUX CONDENSER
D
H
TV
EXPANSION TURBINE
SEPARATOR
G
SEPARATOR
C
TV
F
TV
B
LIQUID FEED SUB-COOLER
DISTILLATION COLUMN
FEED PRECOOLER
N_2 VENT
A
REBOILER
AFTERCOOLER
METHANE PRODUCT
WATER
PRODUCT COMPRESSOR
PRODUCT VAPORIZER

Figure 2.1.1.-61. MODIFIED AMMONIA-ETHYLENE-METHANE CASCADE PROCESS FOR METHANE-NITROGEN SEPARATION

cycle. The process indicated in Figure 2.1.1.-60 is for either a low- or high-pressure natural gas supply, while the cycle in Figure 2.1.1.-61 is applicable only to a high-pressure source.

If the natural gas supply is at a sufficiently high pressure, it is passed through the feed expansion turbine (Figure 2.1.1.-60) where its pressure is reduced to column operating pressure. A low-pressure source would feed directly to the system. The natural gas is cooled in the Feed Precooler B by product methane and exhaust nitrogen vapor. The feed gas passes from the precooler to the feed condenser, where the latent heat of vaporization of recirculated methane supplies the refrigeration necessary to completely liquefy the feed gas. The liquefied feed passes from the condenser to the subcooler, where the vaporization of liquid methane recirculated at a lower pressure subcools the feed stream to the column feed temperature, and then into the column. If the column is operating at higher than the natural gas feed pressure, the liquid is pumped by F into the column.

The gaseous nitrogen from the separator passes through a coil in the liquid feed subcooler to Expansion Turbine G, and then through a second coil in the reflux condenser. From the reflux condenser the nitrogen enters Expansion Turbine H, where the pressure is reduced to near atmospheric. The cooled nitrogen is passed in series through the reflux condenser, liquid feed subcooler, feed condenser, and feed precooler to the atmosphere.

The compressed recirculating methane is condensed in coils in the product vaporizer and the column reboiler. The saturated liquid methane is throttled through Valve J into High-Pressure Separator K and forms a vapor and a liquid phase at a lower temperature.

The first separator pressure is adjusted so the vaporization of the liquid methane and the warming of the gaseous methane in the feed condenser will furnish sufficient refrigeration to completely condense the cooled feed gas. A small stream of the saturated liquid is withdrawn from Separator K to furnish the cooling requirements of the liquid feed subcooler and the reflux condenser, and is expanded through Valve L into Medium-Pressure Separator M. The medium-pressure methane vapor and a portion of the liquid are passed through the liquid feed subcooler.

The liquid withdrawn from the second separator, M, is expanded through Valve N and forms a low-pressure, low-temperature two-phase stream that furnishes the requirements of the reflux condenser that are in excess of the refrigeration recovered from the nitrogen overhead. The low-pressure methane vapor from the reflux condenser passes through the liquid subcooler and the feed condenser, is compressed to the operating pressure of Separator K, and joins the main methane stream at R. The medium-pressure methane vapor from the liquid feed subcooler passes through the feed condenser, is compressed to the operating pressure of Separator K, and also joins the main methane stream at R. The vaporized high-pressure methane stream from the feed condenser passes through Heat Exchanger Q, where it cools the compressed methane vapor, and joins the other streams at R. The combined methane stream is compressed sufficiently to permit complete condensation in the column reboiler and product vaporizer. The compressed methane is cooled in the ethylene evaporator and passed through Heat Exchanger Q to the product vaporizer and column reboiler.

The ethylene and ammonia cycles are the same as those of the ammonia-ethylene-nitrogen process.

If the feed is at a sufficiently high pressure, it is passed through the feed expansion turbine. But if it is at low pressure, it is compressed in the feed compressor to column operating pressure. The natural gas is then cooled by countercurrent heat exchange with the product methane and exhaust nitrogen in the feed precooler. The cooled feed gas is partially condensed and fed to a plate near the bottom of the column in a 2-phase stream (D, Figure 2.1.1.-60). The product methane from the bottom of the column passes through the vaporizer and the feed precooler, and discharges to the pipeline through the product compressor and aftercooler. The overhead nitrogen is cooled in Expansion Turbine G, passed through the reflux condenser, and again cooled by expansion through Turbine H. The low-pressure nitrogen stream passes through the reflux condenser, feed condenser, and feed precooler to the atmosphere.

The recirculating methane condensed in the product vaporizer and column reboiler is throttled to a medium pressure to supply the feed condenser requirements. A portion of the liquid formed is throttled to furnish the requirements of the reflux condenser that are in excess of the refrigeration recovered from the overhead nitrogen. The remainder of the cycle is similar to the system for subcooled liquid feed.

The system can be modified for high-pressure natural gas supply and subcooled liquid feed to the column. The gas enters the feed-expansion turbine (Figure 2.1.1.-61) and is reduced to a pressure that will permit complete condensation at the product vaporizer temperature. The saturated liquid feed is subcooled by evaporating liquid methane and is throttled through Valve C into the column. The product methane, recirculated methane, and overhead nitrogen streams are treated the same as in the low-pressure feed cycles.

The heat load on the liquid feed subcooler and the reflux condenser is determined by the condition of the feed entering the column. For partially condensed feed to the column, the subcooler requirements would be decreased, and the reflux condenser load increased. The cycle would remain unchanged.

The cascade cycles are utilized to provide a low-temperature heat sink for removing the heat of compression from the recirculating fluids. This permits the nitrogen and methane compression to be carried out at a high density (low temperature), which results in reduced compressor energy requirements. However, when compressor efficiency is introduced to the calculations, the cascade requirements are increased until a compression cycle for a single fluid becomes most efficient, rejecting heat at atmospheric temperature, at which compressor aftercooling is most economically available by proper heat exchange between the compressor inlet and the aftercooler discharge streams. By this method the useful refrigeration in the cold low-pressure methane is employed to cool the compressed methane stream.

In the cycle presented in Figure 2.1.1.-62, the reduced pressure recirculating methane streams are warmed to atmospheric temperature by countercurrent heat exchange with the compressed recirculating methane. The low- and medium-pressure methane streams are compressed to the pressure of the first methane separator and aftercooled before recombination with the higher pressure stream. The final compression is to a pressure that will permit complete liquefaction at column reboiler temperature. The remainder of the methane cycle is the same as with the cascade systems.

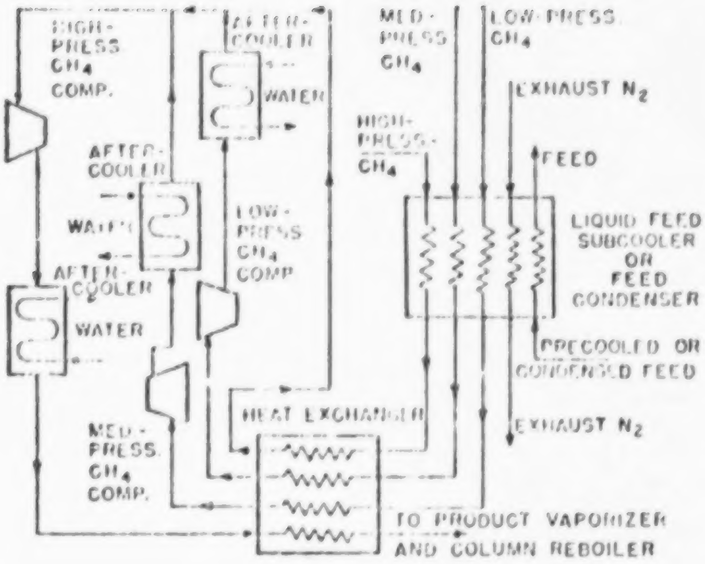

HIGH-PRESS CH4 COMP.
AFTER-COOLER
WATER
AFTER-COOLER
WATER
AFTER-COOLER
WATER
LOW-PRESS CH4 COMP
MED-PRESS CH4 COMP.
HEAT EXCHANGER
MED-PRESS CH4
LOW-PRESS CH4
EXHAUST N2
HIGH-PRESS-CH4
FEED
LIQUID FEED SUBCOOLER OR FEED CONDENSER
PRECOOLED OR CONDENSED FEED
EXHAUST N2
TO PRODUCT VAPORIZER AND COLUMN REBOILER

Figure 2.1.1.-62. RECIRCULATING METHANE PROCESS FOR METHANE-NITROGEN SEPARATION

The four processes described are illustrations of those proposed for nitrogen separation. For the conditions studied, the calculated requirements indicate that the methane cycle is the most efficient process for rectifying high-nitrogen-content methane-nitrogen mixtures. For a high-pressure feed the modified methane cycle would have the lowest energy requirements; these are minimized when the column operating pressure is maximized against construction cost and the column feed is partially condensed. The higher the nitrogen content, the larger the portion of feed that should be condensed for a minimum overall column energy input. For 1000-psia feed gas pressure, 300-psia column operating pressure, and partially condensed feed to the column, the methane cycle compressors require only 0.56% of the total energy (heat of combustion) in the entering gas to effect the separation.

2.1.1.4.3. Investment and Operating Costs

The nitrogen removal processes described have not been subjected to any detailed investment or operating cost analyses. However, the energy required to effect the separation has been computed.

The assumed operating conditions employed in the calculations were —

1. Feed gas composition, 70% methane, 30% nitrogen.
2. Feed gas received at 200 or 1000 psia and 100° F.
3. Product gas at least 97% methane, and separated nitrogen not more than 3% methane. To facilitate calculation, the thermodynamic properties of the two streams are assumed to be those of the pure components.
4. Product methane to be delivered at 800-psia pipeline pressure and 100° F.
5. Temperature difference between countercurrent streams in the heat exchangers to be 3° F or more at all points in each exchanger.
6. Expansion turbines and compressors calculated at 78% adiabatic efficiency.
7. Energy values calculated as shaft horsepower required by the compressors, recovered with expansion turbines.
8. Energy developed in expansion turbines to be recovered in electric generators or direct-coupled compressors. This energy deducted from process compressor requirements to give the net work.
9. Column operated at 1.05 times minimum reflux ratio.
10. Energy and cooling water requirements for nitrogen separation reduced by deducting similar needs for compression of feed from 200 psia to pipeline pressure (800 psia).

The product composition of 97+ % methane is equivalent to removal of 92.8% of the nitrogen present in the feed gas. The 3% methane content in the nitrogen exhaust represents a loss of 1.3% of the methane in the feed. If the overhead stream composition is 99% nitrogen, the methane loss will be reduced to 0.4 of that entering in the feed gas. Since the thermodynamic properties of the separated streams differ slightly from those of the pure components, the listed product compression requirement will be high and the exhaust nitrogen refrigeration yields will be low by approximately 2%. For precise design calculations it will be necessary to construct Mollier diagrams for the terminal compositions.

The 78% efficiency used in calculating compressor requirements and expansion turbine recoveries represents an efficiency attainable by modern centrifugal equipment. Heat losses from the column and exchangers, and pressure drops in piping and equipment are functions of the physical layout of plants, and are, therefore, not considered here.

A summary of the net compression energy and cooling water requirements for the various processes considered and for the different cycles as influenced by feed pressure, column feed conditions, and operating pressure is present in Table 2. 1. 1. -7. The compressor energy and cooling water required to increase the feed gas pressure from 200 to 800 psia for delivery to the pipeline, which would be necessary if the separation plant were not in operation, are given at the end of the table.

Table 2. 1. 1. -7. COMPRESSOR ENERGY AND COOLING WATER REQUIREMENTS, AS INFLUENCED BY COLUMN FEED COMPOSITION, COLUMN PRESSURE, AND FEED GAS PRESSURE, OF FIVE INDEPENDENT CYCLES IN THE SEPARATION OF A 70% METHANE-30% NITROGEN MIXTURE

Basis: Product composition ≥ 97% CH_4; nitrogen exhaust composition ≤ 3% CH_4; product delivered to pipeline at 800 psia and 100°F

				NET REQUIREMENTS			
				Per lb mole Feed Gas		Per MCF Feed Gas	
Refrigeration Cycle	Column Feed Condition	Column Pressure, psia	Feed Gas Pressure, psia	Compressor, Btu	Cooling Water, Btu	Compressor, Btu	Cooling Water, Btu
N_2 Expansion (Cost's)	Saturated liquid	200	200	6746	7435	17,803	19,621
		200	1000	4786	5219	12,630	13,773
NH_3-C_2H_4-N_2	Saturated liquid	200	200	7765	7283	20,492	19,220
		200	1000	5806	5067	15,322	13,372
NH_3-C_2H_4-CH_4	Subcooled liquid	200	200	5192	4952	13,702	13,068
		200	1000	4134	3641	10,910	9,609
		300	200	5012	4563	13,227	12,042
		300	1000	3954	3252	10,435	8,582
Modified NH_3-C_2H_4-CH_4	Subcooled liquid	200	1000	2894	2776	7,637	7,326
NH_3-C_2H_4-CH_4	Partially-condensed	200	200	3265	3133	8,616	8,268
		200	1000	2207	1822	5,824	4,808
		300	200	3175	2964	8,379	7,822
		300	1000	1744	1280	4,602	3,378
CH_4 cycle	Partially-condensed	200	200	3044	3136	8,033	8,276
		200	1000	1986	1825	5,241	4,816
		300	200	2785	2830	7,350	7,468
		300	1000	1354	1146	3,573	3,024
Compressing feed gas from 200 to 800 psia				2088	2262	5,510	5,967

2.1.2. Liquefaction Cycle

In natural gas liquefaction cycles energy, in the form of sensible and latent heat, is removed. There are two widely used methods for heat removal.

1. The cascade cycle in which heat is transferred through refrigerants to a high-level heat sink, such as cooling water or air.
2. The expander cycle, in which cooling is accomplished as the gas does work in an expansion engine.

Liquefaction cycles can use either or both of these concepts.

2.1.2.1. Cascade Cycle

2.1.2.1.1. Process Description

To liquefy a gas having a dew point below ambient temperature, heat must be removed from the gas. Heat is removed from the gas at a low temperature and transferred to an available heat sink (cooling water or air) at high temperature.

In a continuous system, a cyclic process must be used to allow uninterrupted operation. The energy balance around such a refrigeration system (Figure 2.1.2.-1) can be expressed according to the first law of thermodynamics as:

where $$Q = Q_0 + L$$

Q = the heat rejected into the cooling water

Q_0 = the heat extracted from the surroundings

and L = the work required by the system

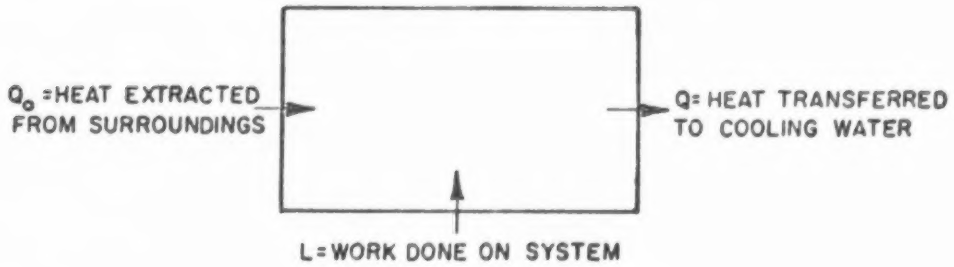

Q_0 =HEAT EXTRACTED FROM SURROUNDINGS
Q= HEAT TRANSFERRED TO COOLING WATER
L=WORK DONE ON SYSTEM

Figure 2.1.2.-1. ENERGY BALANCE OF REFRIGERATION SYSTEMS[44]

An ideal cycle for transferring heat from a low temperature to a higher temperature would use only reversible operations. Such a cycle is in the T-S diagram of Figure 2.1.2.-2.

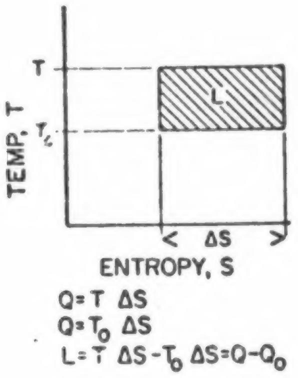


Figure 2.1.2.-2. T-S DIAGRAM SHOWING IDEAL CYCLE OF REFRIGERATION SYSTEM [44]

For a refrigeration cycle it is important to know how much refrigeration (Q_0) can be produced by the work (L) required for the operation of the cycle. For the ideal process, this ratio can be directly derived from the T-S diagram, namely:

$$\frac{Q_0}{L} = \frac{\Delta S \cdot T_0}{\Delta S \cdot (T - T_0)} = \frac{T_0}{T - T_0}$$

The work (L) is the minimum work required. An actual cycle will require more work than the ideal cycle because of many irreversible changes which will occur in the system. The additional work is caused by the irreversibility of the operation, which can be expressed by the entropy increase ΔS^{irr}. The additional work is then $L^{irr} = T^{o} \cdot \Delta S^{irr}$. Therefore, the more it can approach the irreversibility of the ideal cycle, the more efficient a liquefaction cycle will be.

One of the more frequently used methods for the supply of refrigeration at lower temperatures is evaporation of a liquid refrigerant at a low pressure and temperature. The latent heat required for the evaporation is extracted from the surroundings; that is, refrigeration is supplied to the surroundings. The refrigerant vapor is reliquefied at a higher temperature and pressure, thereby transferring the latent heat to the cooling water (Figures 2.1.2.-3 and 2.1.2.-4). The temperature difference that can be reached by such a cycle is determined by the vapor pressure curve of the refrigerants (Figure 2.1.2.-5).

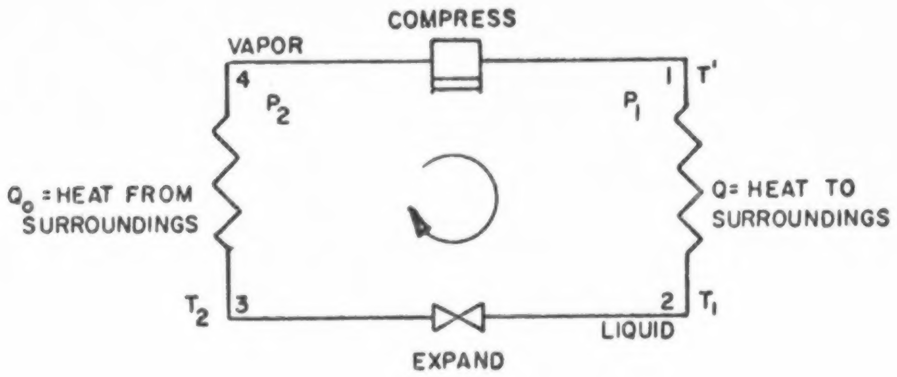

COMPRESS
VAPOR
4
P_2
P_1
1
T'
Q_o = HEAT FROM SURROUNDINGS
Q= HEAT TO SURROUNDINGS
T_2
3
2
T_1
LIQUID
EXPAND

Figure 2.1.2.-3. REFRIGERATION CYCLE USING ONE REFRIGERANT [44]

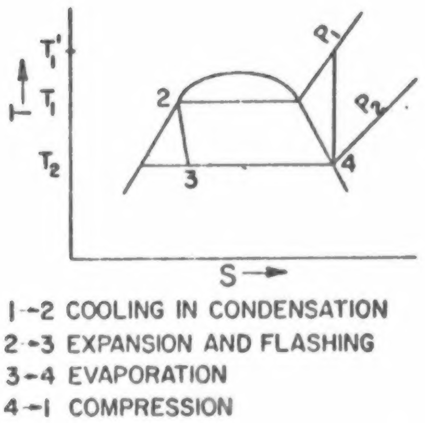

T_1'
T
T_1
T_2
2
3
4
P_1
P_2
S→
1-2 COOLING IN CONDENSATION
2-3 EXPANSION AND FLASHING
3-4 EVAPORATION
4-1 COMPRESSION

Figure 2.1.2.-4. T-S DIAGRAM OF FIGURE 2.1.2.-2. REPRESENTED[44] ON A T-S DIAGRAM

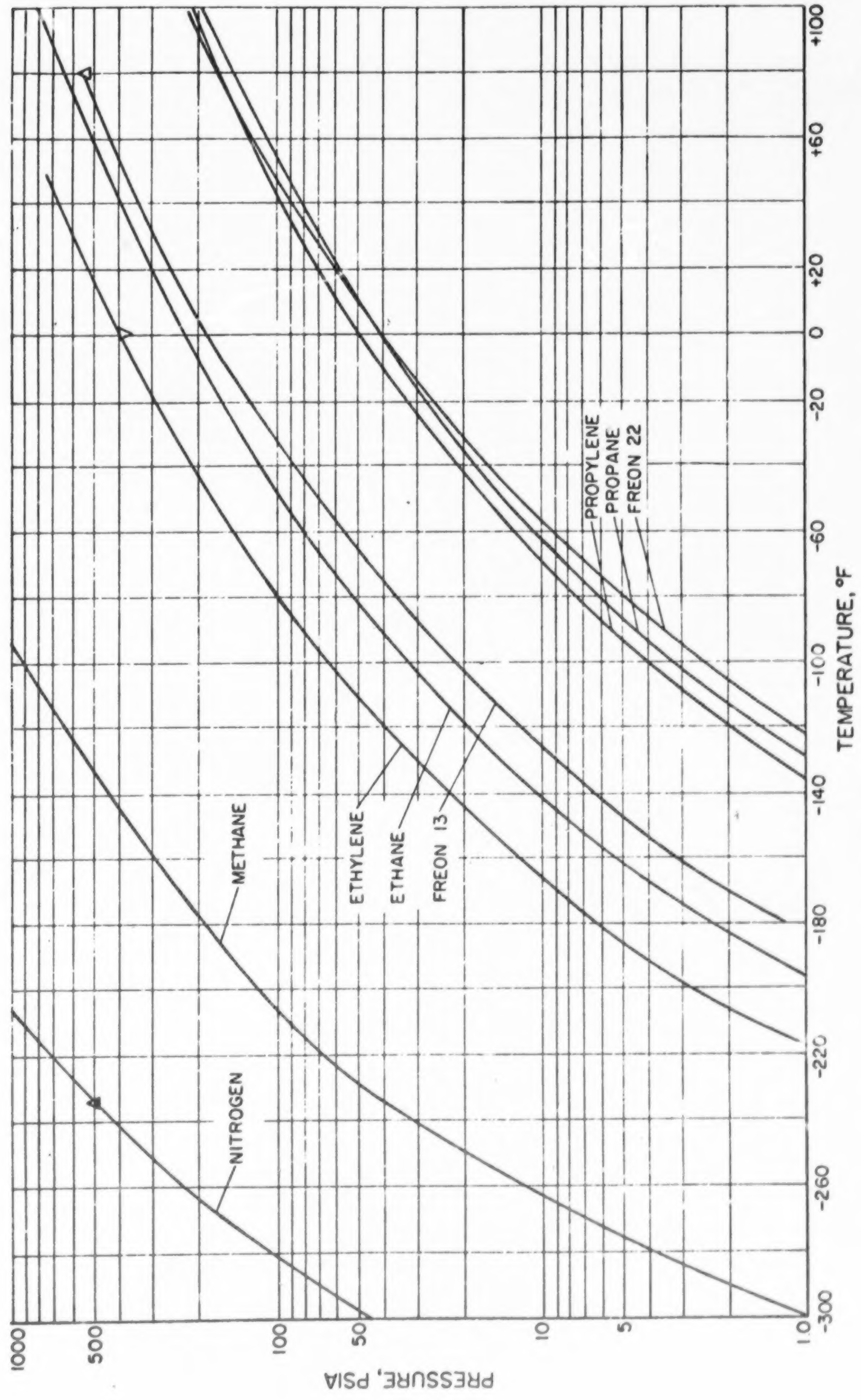


Figure 2.1.2.-5. VAPOR PRESSURES (APPROXIMATE) OF SOME COMMON REFRIGERANTS[44]

To work under realistic conditions the heat of condensation at the condensing pressure should still be appreciable; that is, the condensing pressure should be well below the critical pressure. On the other hand, to limit the size of the vessels and the compression equipment, the evaporating pressure should not be too low.

Considering the vapor pressure curves of gases in the temperature range of +100° to −300°F, it becomes obvious that only a temperature range of approximately 150°F is practically obtainable in a system using one refrigerant. To extend the temperature range to lower temperatures, more than one refrigerant must be used. The lower boiling refrigerant would be chosen by the condition that it liquefy at a reasonable pressure at the evaporating temperature of the higher boiling refrigerant.

A cascade cycle is a system in which a series of refrigerants are used to obtain lower and lower temperatures. The cascade principle was first conceived by Pictet, later carried out by H. Kamerlingh Onnes and W. H. Keesom. [37]

Referring to the vapor pressure curves, it is seen that such a system could be composed of the series propane-ethylene-methane; a simplified cascade system using these three refrigerants is presented in Figure 2.1.2.-6. The cascade cycle shown in Figure 2.1.2.-6 must be considerably improved to become a practical and efficient natural gas liquefaction cycle. The improvements are in principle accomplished by introducing features that reduce the irreversibility of the process. The perfected cascade cycle is shown in Figure 2.1.2.-7.

The natural gas at a pressure of 450-500 psig is combined with the methane recycle from compressor CP-1 and split into two streams. One passes through multifluid countercurrent heat exchanger E-1, where it is cooled against the returning low-pressure methane flash vapors. The other one passes through a series of heat exchangers E-3A, E-3B, and E-2A, where it is precooled by evaporating propane or ethylene. The two streams are then combined and liquefied in condenser CD-1, utilizing refrigeration supplied by evaporating ethylene.

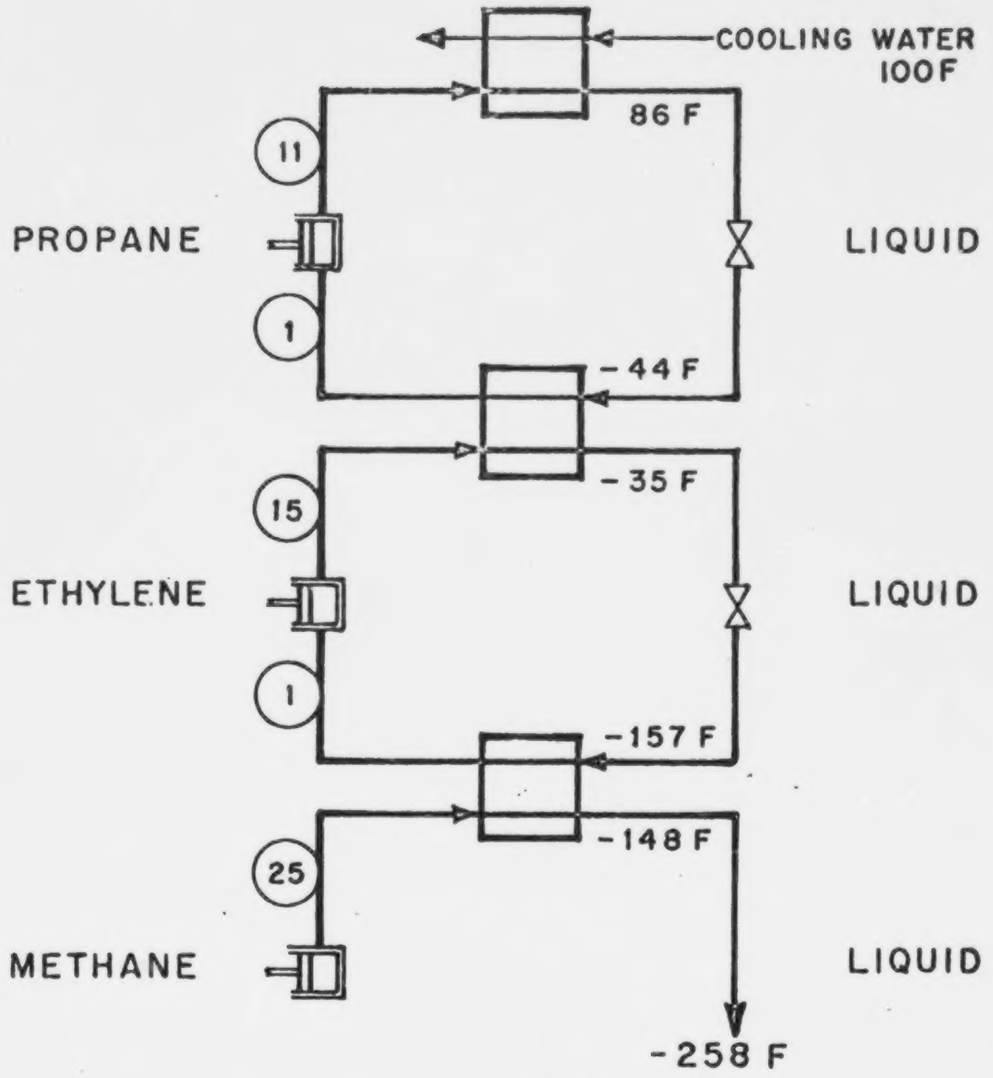

COOLING WATER
100F
86 F
11
PROPANE
LIQUID
1
-44 F
-35 F
15
ETHYLENE
LIQUID
1
-157 F
-148 F
25
METHANE
LIQUID
-258 F

Figure 2.1.2.-6. SIMPLIFIED CASCADE SYSTEM SHOWING PRINCIPLES OF THE METHOD[44]

The liquefied high-pressure natural gas is then expanded into phase separator PS-1 to an interstage pressure of the methane recycle. The flash vapor is warmed up in heat exchanger E-1 and recompressed in the second and third stages of compressor CP-1. The liquid fraction is subcooled in subcooler SC-1 against low-pressure flash vapor and then expanded into phase separator PS-2. The pressure in this separator must be held high enough to allow transfer to the liquid storage. The flash vapor is returned and warmed up in subcooler SC-1 and heat exchanger E-1. Flash vapor from the storage tank, which is normally held at atmospheric pressure, can be recompressed in cold blower CP-4 and then joined with the low-pressure flash vapor from PS-2.

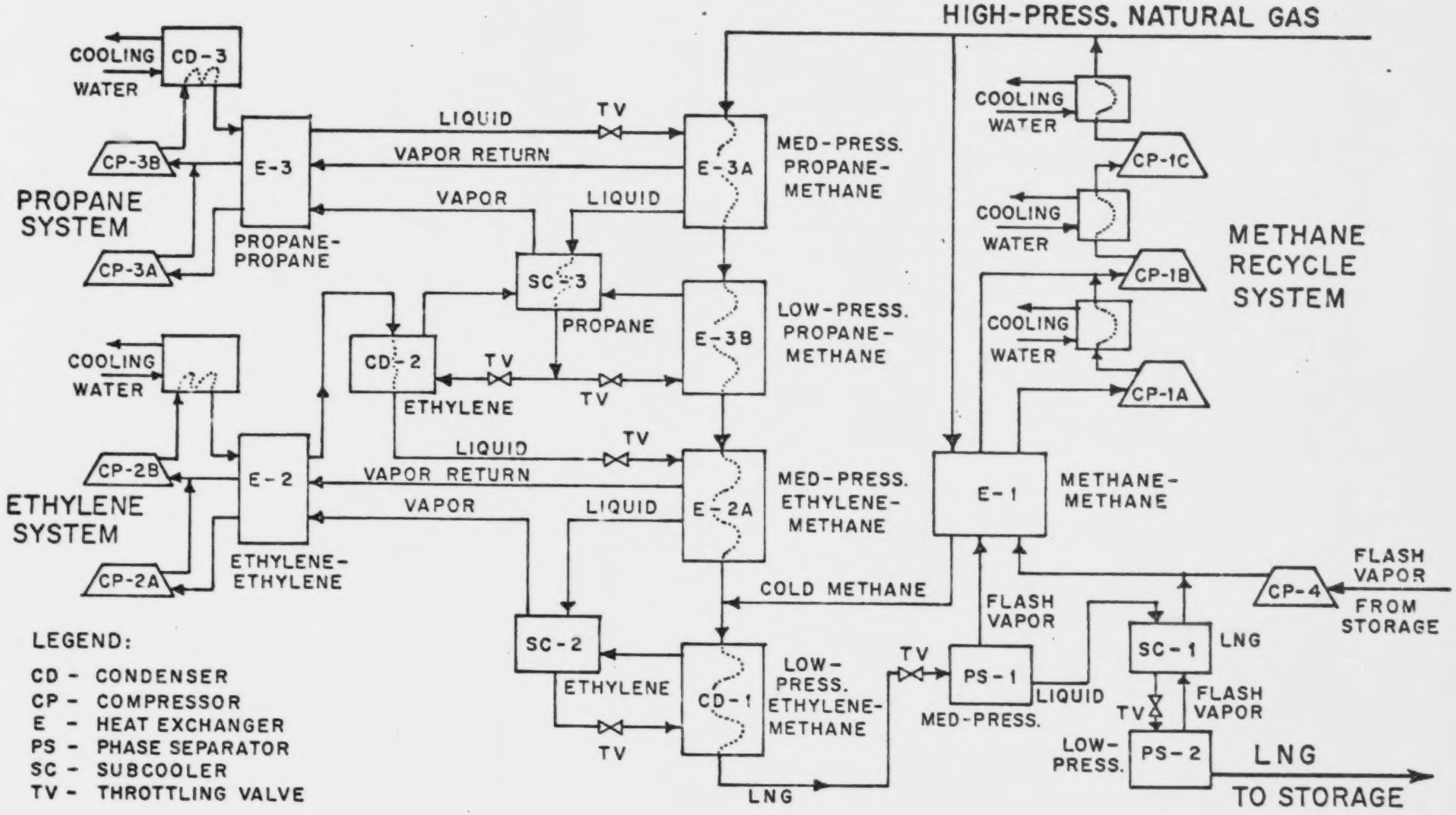

HIGH-PRESS. NATURAL GAS
COOLING
WATER
CD-3
CP-3B
CP-3A
PROPANE SYSTEM
E-3
PROPANE-PROPANE
LIQUID
TV
VAPOR RETURN
VAPOR
LIQUID
E-3A
MED-PRESS. PROPANE-METHANE
SC-3
PROPANE
E-3B
LOW-PRESS. PROPANE-METHANE
CD-2
TV
ETHYLENE
TV
COOLING
WATER
CP-2B
CP-2A
ETHYLENE SYSTEM
E-2
ETHYLENE-ETHYLENE
LIQUID
TV
VAPOR RETURN
VAPOR
LIQUID
E-2A
MED-PRESS. ETHYLENE-METHANE
COLD METHANE
SC-2
ETHYLENE
TV
CD-1
LOW-PRESS. ETHYLENE-METHANE
LNG
TV
PS-1
MED-PRESS.
FLASH VAPOR
LIQUID
E-1
METHANE-METHANE
COOLING
WATER
CP-1C
COOLING
WATER
CP-1B
COOLING
WATER
CP-1A
METHANE RECYCLE SYSTEM
CP-4
FLASH VAPOR
FROM STORAGE
SC-1
LNG
FLASH VAPOR
TV
PS-2
LOW-PRESS.
LNG
TO STORAGE
LEGEND:
CD - CONDENSER
CP - COMPRESSOR
E - HEAT EXCHANGER
PS - PHASE SEPARATOR
SC - SUBCOOLER
TV - THROTTLING VALVE

Figure 2.1.2.-7. DETAILED CASCADE CYCLE [44]

Propane returning from the liquefier is compressed in two stages of compressor CP-3 to approximately 180 psig and then condensed by cooling water in condenser CD-3. The liquefied propane is then subcooled in heat exchanger E-3 before it is expanded into the shell side of heat exchanger E-3A. The flash vapor and the vapor generated by the cooling of the natural gas are returned to the interstage of compressor CP-3B after having been warmed up in heat exchanger E-3.

Liquid propane is withdrawn from the shell side of E-3A, and subcooled in subcooler SC-3. One portion is expanded into heat exchanger E-3B, where it evaporates, offering refrigeration for cooling of the natural gas. The other portion is expanded into the ethylene condenser CD-2, where it evaporates. The low-pressure vapors are returned to the section of compressor CP-3A after having been warmed up in subcooler SC-3 and heat exchanger E-3.

The ethylene is compressed in the two stages of compressor CP-2. It is precooled in heat exchanger E-2 against the returning low-pressure vapors, and then liquefied in CD-2. The liquid ethylene is then expanded into heat exchanger E-2A, which operates at the interstage pressure of CP-2. The flash vapor and the vapor generated by cooling of the natural gas are warmed up in heat exchanger E-2, and returned to the interstage of compressor CP-2. Liquid ethylene is withdrawn from the shell side of E-2A, subcooled in subcooler SC-2, and then expanded into the shell side of condenser CD-1. Here the ethylene evaporates, supplying refrigeration for the condensation of the natural gas; the low-pressure vapor is warmed up in subcooler SC-2 and heat exchanger E-2, and then recompressed in the ethylene compressor CP-2.

In comparison to the simplified cycle, the following improvements were made:

1. Multistage compression, reducing the power of compression.
2. Multistage expansion of the liquefied refrigerant, allowing recompression of flash vapors from the various interstage pressures.
3. Heat exchange between entering gas or liquid streams and effluent cold vapors, allowing recovery of the sensible heat contained in the cold flash vapors.

The cascade system can be adapted to any cooling curve; that is, the quantity of refrigeration supplied at the various temperature levels can be chosen so that the temperature differences in the evaporators and heat exchangers approach a practical minimum. (Small temperature differences mean low irreversibility, and therefore low power consumption.)

A cascade liquefaction system consists mainly of the following components:

1. The compressor unit required to perform the work necessary for the transfer of heat from lower to higher temperatures.
2. The heat exchange unit required to transfer heat from the liquefying natural gas to the refrigerants, from the returning cold vapors to the entering refrigerants, and from the lower boiling refrigerant to the higher boiling refrigerant.
3. The cooling unit for the condensation of the highest boiling refrigerant i.e., cooling tower for a cooling-water system or air-fin coolers.
4. The instrumentation and control unit for the operation of the entire system.

2.1.2.1.2. Investment Costs

Cascade plants of both the base-load and peakshaving sizes have been built. The gas liquefaction plant at Arzew, Algeria, uses the classical cascade cycle. Details of its investment were reported by Bourguet and Perret.[3] The plant has three parallel streams, each rated at 50 million SCF/day or a total of 150 million SCF/day. It reached full capacity in March 1965 and has been operating satisfactorily ever since. A 20-year life span is expected based on economical repair resulting from normal operations.

Table 2.1.2.-1 illustrates some of the important physical factors in the erection of this facility. It indicates the magnitude of the effort.

Table 2.1.2.-1. PHYSICAL STATISTICS OF ARZEW LNG PLANT[3]

Capacity, 10^6 SCF/day	150
LNG Storage Capacity, bbl	445,000
Total Area, acres	181
Length of Fence, miles	2.25
Concrete, cubic yards	52,000
Weight of Material and Equipment, tons	26,000
Length of Piping Above 2-in. Diameter, miles	87
Total Heat Exchanger Area, sq ft	105,000
Power Installed, hp	120,000
Flow of Seawater Cooling, gpm	152,000
Job Site Erection Time, man-hours	7,500,000
Design Time, man-hours	500,000
Length of Electric Cable, miles	131

Table 2.1.2.-2 illustrates the original cost factors. During the summer of 1966 certain improvements were made to increase the capacity 30% to 200 million CF/day with an additional cost of only $1,600,000.

Table 2.1.2.-2. OVERALL COST FACTORS FOR ARZEW LNG PLANT

	Value, $\$10^6$
Hassi Er R'Mel Pipeline	70
Hassi Er R'Mel Liquefaction Plant	84 (150 X 10^6 CF/ day)
Port Arzew Dock Facilities	10
Total	164

No breakdown in the cost of the individual compressors, power units, etc., is available, but the equipment is shown schematically in Figure 2.1.2.-7.

After the purified gas stream is split into three equal streams, it enters the liquefaction lines. Each liquefaction line has three steam turbine-driven compressors, one 15,600-hp unit supplying two propane refrigeration stages (medium and low), one 11,200-hp unit supplying three ethylene refrigeration stages, and one 4600-hp compressor supplying three methane refrigeration stages.

Based on the 200 million CF/day uprated capacity and the $85,600,000 cost, the investment cost is $428/1000 CF/day. Reference 3 states that, at 1964 price levels, a 40 million CF/day cascade liquefaction system in the U.S. has a unit cost of $350/1000 CF/day.

Figure 2.1.2.-8 illustrates another analysis[31] that is in substantive agreement with the above; it illustrates that unit cost of a liquefaction plant varies somewhat with size. For example, two 200 million CF/day plants cost approximately 17% more than one 400 million CF/ day plant, which would mean that large liquefaction complexes can economically increase their capacity in steps rather than investing a large amount initially in the hope of using the extra capacity in the future. Since these data were compiled in May 1965, plant costs have probably increased 10%

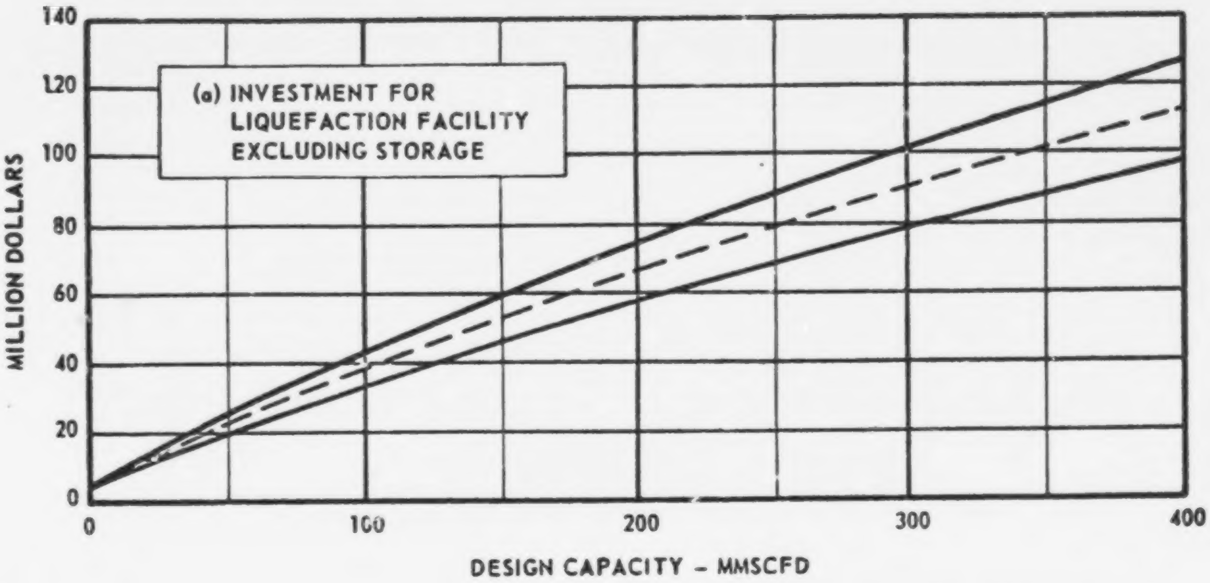


Figure 2.1.2.-8. INVESTMENT FOR LIQUEFACTION FACILITY EXCLUDING STORAGE[31]

2.1.2.1.3. Operating Costs

Operating costs include all costs related to liquefaction from the time the gas enters the plant gas compressor until it is pumped into the storage tank. Two examples of the computation method will be illustrated.

For the first example we will use the Arzew plant, which has a capital cost for liquefaction of $85,600,000. We may assume a 15-year straight-line depreciation with 6% interest rate and a 6% combined insurance and tax rate. These figures (principal, interest, insurance, and tax) will make up the capital cost.

The plant will be assumed to operate 300 days/yr at its full capacity of 200 million CF/day. While it is theoretically possible to operate for 330 days or more per year, experience indicates that some allowance should be made for political as well as mechanical unrest. The gas will be assumed to have a heating value of 1100 Btu/CF and to cost 6¢/million Btu. The cost of fuel gas energy will be taken as 15% of the cost of the sale gas — the sale gas being the natural gas to be liquefied. The operating expense will consist of supervisory and direct labor in connection with liquefaction as well as chemicals, supplies, spare parts, maintenance labor, and other factors. Based on cryogenic plants and remote operation this will be taken at 5% of the invested capital.

Table 2.1.2.-3 summarizes the cost accounting and shows that the cost of LNG delivered to the storage tank is approximately 35.8¢/1000 CF.

Table 2.1.2.-3. CALCULATION OF UNIT COSTS OF AN ANNUAL LNG PRODUCTION OF 60.0 BILLION CF

1. Liquefaction Facility Investment	$82,600,000
2. Capital Cost	
Annual Principal	$ 5,707,000
6% Average Interest	$ 2,568,000
6% Tax Rate and Insurance	$ 5,136,000
	$13,411,000
3. Sale Gas	
(5.45¢/1000 CF) X (60 X 10^9 CF)	$ 3,270,000
4. Fuel Gas	$ 490,000
5. Annual Operating Expense	
5% X $85,600,000	$ 4,280,000
6. Total Cost	$21,451,000
7. Unit Cost of LNG	35.8¢/1000 CF

In computing the total cost of delivered LNG, the cost of shipping and the cost of storage at the sending and receiving terminal are the largest additional factors. It is obvious that the capital cost is by far the largest factor in the expense. Item 5 in the table may, at first glance, appear high; this figure has been estimated liberally due to the significant extra costs involved in successfully operating sophisticated equipment in foreign countries and in remote areas in those places. It is also worth noting that fuel gas, or liquefaction cycle efficiency, is of minor importance in the overall economics.

Another approach is illustrated by Table 2.1.2.-4.

Table 2.1.2.-4. CALCULATION OF UNIT COSTS[2]
(Annual exship LNG delivery: 50 X 10^{12} Btu)

Liquefaction Facility Investment, $\$10^3$	60,000	
Purchased Fuel Gas at \$0.06/ 10^6 Btu, $\$10^3$/yr	370	
Purchased Sale Gas at \$0.06/ 10^6 Btu, $\$10^3$/yr	3,132	
Other Operating Expenses, $\$10^3$/yr	3,500	
Annual Capital Charges Cum Taxes at 21 and 16% of Initial Investment, $\$10^3$/yr	12,600	9,600
Total Annual Revenue to Liquefaction Company, $\$10^3$/yr	19,602	16,602
Total Annual Sales by Liquefaction Company, 10^9 Btu	52,200	
Unit Price f.o.b. Tankers, $\$10^6$ Btu	0.375	0.318

The assumed heating value of the piped gas, 1100 Btu/CF, and the cost of gas, \$0.06/million Btu (5.45¢/1000 CF), is the same as in Table 2.1.2.-3. The plant in Table 2.1.2.-4 will actually need to liquefy approximately 5% more than 50 trillion Btu to cover the losses of shipping and handling. This 160 million CF/day facility has a liquefaction investment of \$60 million, which is confirmed by Figure 2.1.2.-8.

The unit costs of 28.9¢/1000 CF and 34¢/1000 CF, as shown in Table 2.1.2.-4 are somewhat lower than 35.8¢/1000 CF, but the latter figure includes a more generous safety factor for capacity reduction: 300 days per year vs. 345 days for Table 2.1.2.-4. Although there is a difference in production capacity, 160 million CF/day in Table 2.1.2.-4 and 200 million CF/day in Table 2.1.2.-3 this scale factor is not appreciable. However, because of the capital intensiveness of a liquefaction plant, it would probably be uneconomical to operate the large plant at 50% of rated capacity for prolonged periods.

The sizes selected for the two examples compare as follows to the two base-load plants now in construction: Esso Libya, 422 million CF/day of 1100 Btu/CF (actual LNG will be 345 million CF/day of 1350 Btu/CF product), and the Phillips-Marathon Alaskan Project rated at 139 million CF/day (approximately 1000 Btu/CF gas).

The unit costs outlined above indicate a rather high LNG cost. Considerable work is now being done to improve the capital structure; this work will be described in Section 2.1.2.2.

2.1.2.2. Modified Cascade Cycle

2.1.2.2.1. Process Description

The modified cascade cycle overcomes some of the disadvantages of the conventional cascade cycle. In this modification, as shown in Figure 2.1.2.-9, the refrigerants are combined into a common stream. The essential feature of this cycle is the progressive cooling and condensation of the gas obtained by successive cooling steps with one refrigerant. The refrigerant consists of a gas mixture, principally ethane and LPG, extracted from the inlet gas stream that is enriched by nitrogen from an outside source, in order to compensate for nitrogen losses. Composition of the cycle gas is controlled to obtain the required refrigeration at each level of the cascade. The control is accomplished by monitoring the cycle gas on a continuous basis by the use of a gas chromatograph.

Use of the incorporated cascade cycle permits the utilization of a simpler piping system and fewer compressor cylinders than could be achieved with a standard cascade cycle. Since there is only one compressor, the simplified control of the cycle to take care of varying operating conditions such as the composition of the natural gas, cooling water temperature, and gas pressure will readily modify the composition of the cycle coolant to maintain operation at optimum efficiency. The total amount of refrigeration is always available in the different sections of the cycle. For example, the refrigeration is distributed to any section of the cycle which may require more or less refrigeration during changes in composition of the inlet gas stream.

A patent by Roche, assigned to S.E.G.A.N.S., Paris, embodies some of these modifications and also enables the separation and regulation of the so-called neutral gases such as argon, nitrogen, and helium during the liquefaction process.

Natural gas at pressures ranging from 550 to 750 psig is cooled and liquefied in the tube side of exchangers C_3, C_2, and C_1. The recycle gas stream is compressed in a reciprocating gas compressor, cooled, and partially liquefied in an aftercooler before entering a separator where a propane-rich phase is obtained. The liquid is throttled through a valve, mixed with the return recycle gas from exchanger C_2 and vaporized by passage through the shell side of exchanger C_3. This refrigeration allows for

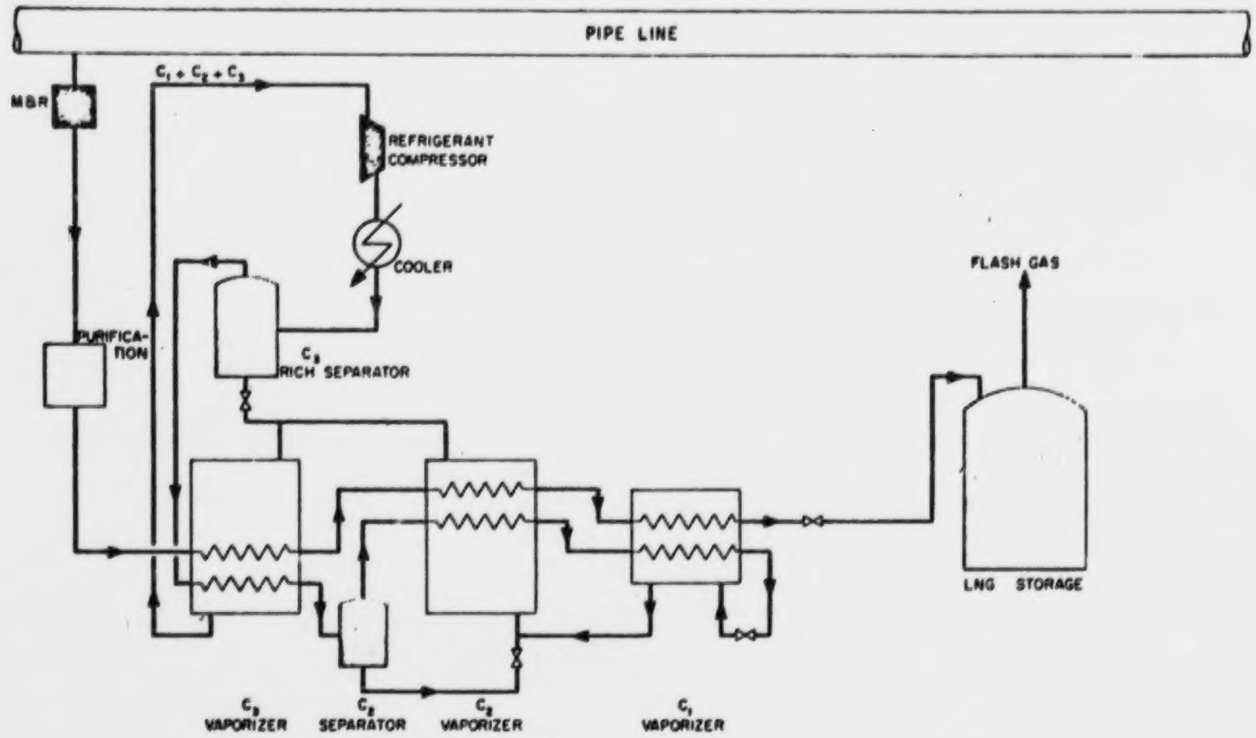

PIPE LINE
MBR
$C_1 + C_2 + C_3$
REFRIGERANT COMPRESSOR
COOLER
PURIFICATION
C_3 RICH SEPARATOR
FLASH GAS
LNG STORAGE
C_3 VAPORIZER
C_2 SEPARATOR
C_2 VAPORIZER
C_1 VAPORIZER

Figure 2.1.2.-9. INCORPORATED CASCADE CYCLE FOR NATURAL GAS LIQUEFACTION [75]

the partial liquefaction of both the natural gas and the recycle gas streams passing through the tube side of exchanger C_3. Vapor from the propane separator is partially condensed in exchanger C_3 after which it enters the ethane separator in an ethane-rich liquid phase. This liquid stream is throttled through a valve, mixed with the return recycle gas from exchanger C_1 and vaporized by passage through the shell side of exchanger C_2, allowing the liquefaction of the natural gas to proceed while the recycle gas is partially condensed. This liquid mixture enters a third separator (not shown), where a methane-rich liquid phase is obtained. The liquid is throttled through a valve, mixed with the returning recycle gas stream from a subcooler (not shown) and vaporized in passing through the shell side of exchanger C_1. Subcooling of the natural gas prior to transfer to storage is accomplished by adjustment of the recycle gas composition.

2.1.2.2.2. Investment Costs

Although no base-load modified cascade plants are in use, a Canadian peakshaving plant has recently started operation and a 1 million CF/day installation in Nantes, France, has been in operation since September 1964. The proponents of this method claim it is especially well suited to large base-load liquefaction schemes because —

a. It replaces the cascade cycle's multistage centrifugal compressors, many of which operate at low temperatures, with fewer, more efficient axial compressors operating at ambient temperatures.

b. It replaces the three refrigerant cascade pure fluids with one single fluid, thus reducing the piping and storage costs.

c. It has fewer, though larger, heat exchangers.

d. It is easier to control.

It has been estimated[57] for the Arzew, Algeria, cascade plant that the use of two parallel modified cascade lines of 100 million CF/day each instead of the three parallel lines of conventional cascade would have resulted in a capital investment reduction of 40%. Also, it is claimed that the greater simplicity of this method would reduce the operating labor needed, which is a significant expense. On the other hand, an increase in the fuel cost is expected. Reference 1 states that a single 67,000-kW input axial compressor is capable of developing enough refrigeration to liquefy 150 million CF/day. Reduction in capital expense and simplified operation will be especially important for future LNG plants.

Future liquefaction complexes will be capable of continuous capacities of 750 million CF/day and more. But, unless these complexes have the simplest possible liquefaction systems, the huge capital requirements will discourage investment.

2.1.2.2.3. Operating Costs

Having analyzed the operating cost of the cascade we can estimate the cost of the modified cascade by reducing the capital investment 40% and making proportional changes in the other items.

The fuel cost increase can be estimated by noting that a 67,000-kW turbine is required to obtain 150 million CF/day, which is equivalent to 0.25 kWhr/lb of LNG. The conventional cascade handling an identical composition gas stream requires 0.21 kWhr/lb of LNG.

Table 2.1.2.-5shows the computation. Note that 24.1¢/ 1000 CF is approximately 67.5% of the cost of the conventional cascade. Some futher benefits can be expected with larger size units; for approximately double this size plant, the unit cost would be reduced approximately 17%.

Table 2.1.2.-5. CALCULATION OF UNIT COSTS FOR AN ANNUAL LNG PRODUCTION OF 60 BILLION CF

1. Liquefaction Facility Investment	
0.60 + ($85.6 X 10^6)(See Table 2.1.-3)	$51,360,000
2. Capital Cost	
Annual Principal	$ 3,424,000
6% Average Interest	$ 1,541,000
6% Tax Rate and Insurance	$ 3,082,000
	$ 8,047,000
3. Sale Gas (5.45¢/1000 CF) X(60 X 10^9 CF)	$ 3,270,000
4. Fuel Gas	$ 583,000
($490,000) X(0.25/0.21 kWhr/lb)	
5. Annual Operating Expense	
5% of $51,360,000	$ 2,568,000
6. Total Cost	$14,468,000
7. Unit Cost of LNG	24.1¢/1000 CF

2.1.2.3. Expansion Cycle

2.1.2.3.1. Process Description

A considerable amount of commercial low-temperature refrigeration is produced by cycles employing expansion turbines or engines which expand gas while performing external work. Most large-tonnage air separation plants employ expander cycles of one form or another. However, the application of expander cycles to natural gas liquefaction is not anticipated, except for small plants and in situations where it is possible to utilize considerable amounts of energy which are normally being wasted by letting natural gas expand through valves. This is due to the higher power requirements normally required for expander cycles as compared to cascade cycles.

A simple expander cycle as proposed by Claude is shown in Figure 2.1.2.-10. Treated feed gas, along with recycle flash gas, is compressed and flows to a heat exchanger. A portion of the high-pressure gas is withdrawn at an intermediate point in the exchanger, and expanded through either a turbine or engine to produce a cooling effect. The expanded cold gas passes to the cold end of the heat exchanger and is warmed before being recycled to the compressor. The high-pressure gas remaining after the withdrawal of expander feed is further cooled by the expander effluent, and finally expanded through a valve. Liquid leaving the valve is separated from flash vapors and discharged to storage, while the flash vapors are warmed in the heat exchangers and recycled through the compressor.

External work is performed by the expander, either in the form of electrical energy, compressed gas, or pumped liquid, depending on the method of loading the expander.

The simple Claude cycle is sketched on a T-S diagram in Figure 2.1.2.-11. High-pressure gas is cooled along Curve 1-2 at constant pressure. Part of the cooled gas is expanded ideally at constant entropy through a turbine or engine. The remainder of the high-pressure gas is cooled further along line 2-3 at constant pressure before being expanded adiabatically along line 3-4 to form a vapor-liquid mixture. The liquid represented as point 5 flows to storage, while the gas at point 6 is joined with the expander discharge and warmed at constant pressure along line 6-7. Superheated vapor from storage at a lower pressure is warmed along line 8-9.

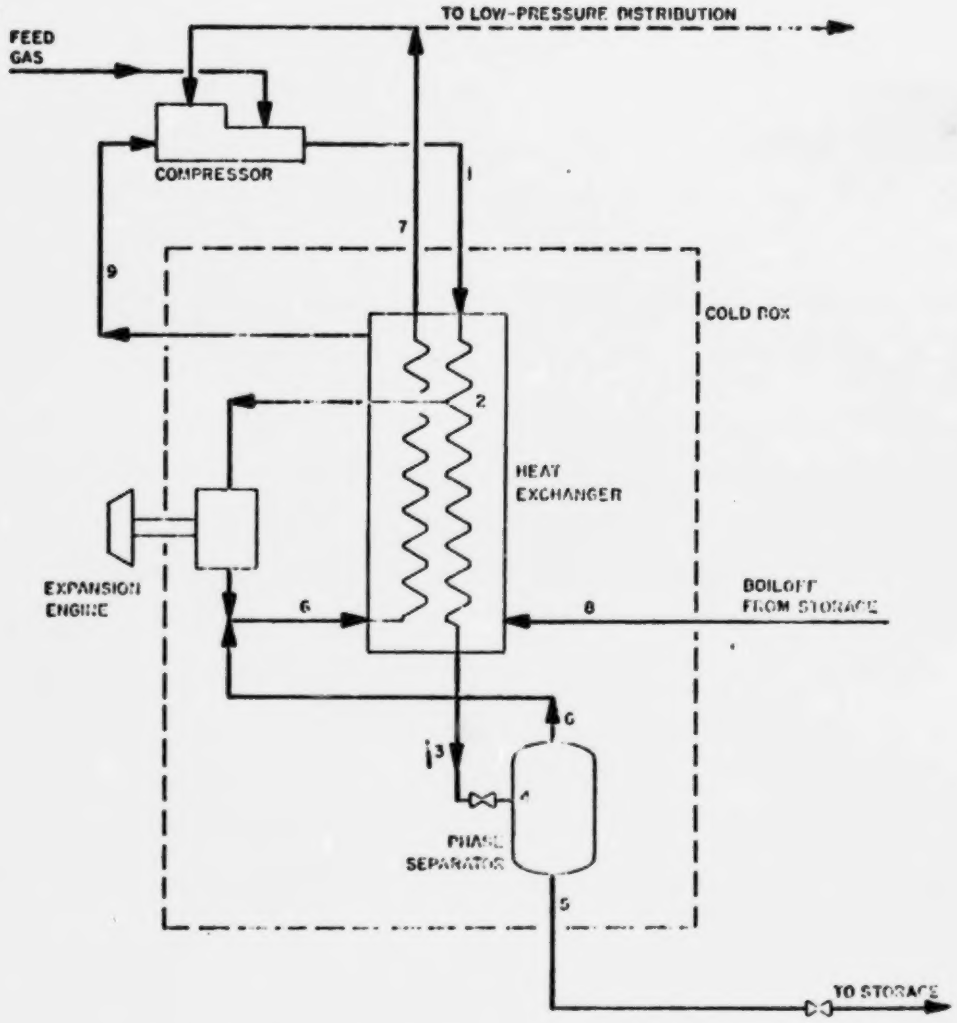

TO LOW-PRESSURE DISTRIBUTION
FEED GAS
COMPRESSOR
1
7
9
COLD BOX
2
HEAT EXCHANGER
EXPANSION ENGINE
6
8
BOILOFF FROM STORAGE
6
3
4
PHASE SEPARATOR
5
TO STORAGE

Figure 2.1.2.-10. SIMPLE EXPANSION ENGINE LIQUEFACTION CYCLE[44]

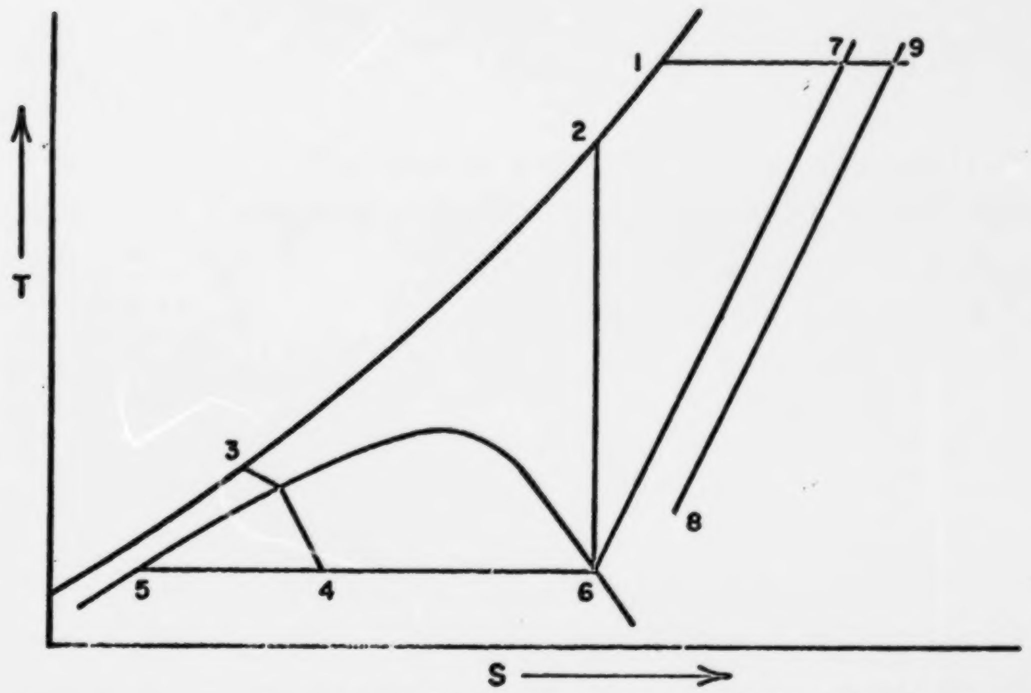


Figure 2.1.2.-11. T-S DIAGRAM OF THE SIMPLE CLAUDE CYCLE[44]

The cooling curve showing the temperature differences existing in the heat exchangers is given in Figure 2.1.2.-12.

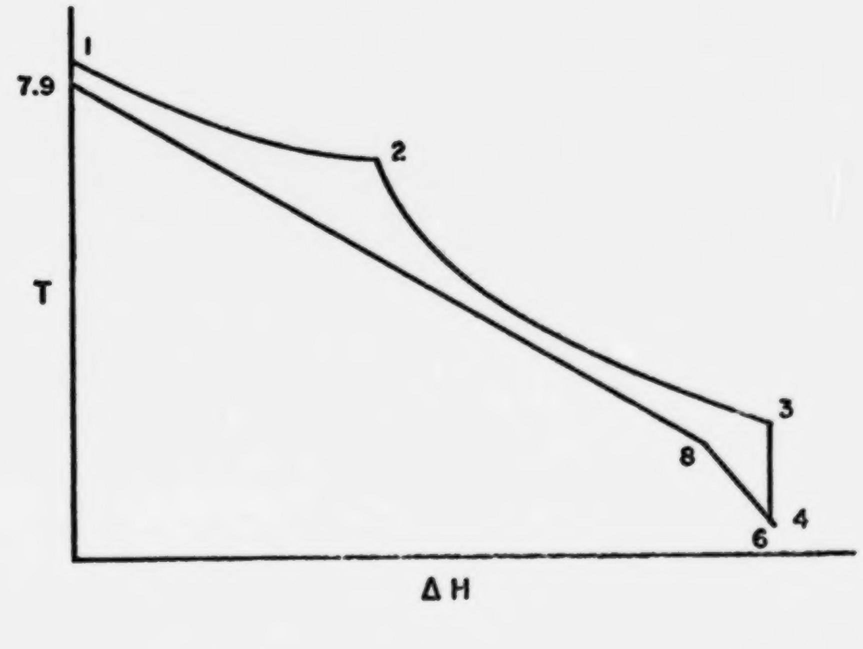


Figure 2.1.2.-12. TEMPERATURE DIFFERENCES IN THE HEAT EXCHANGERS IN THE CLAUDE CYCLE[44]

Since the streams being warmed in the heat exchangers are usually at low pressure, the specific heat is relatively constant over the range of temperatures existing in the exchangers. Therefore, in order to minimize the temperature differences in the exchangers, it is important to select a pressure of the stream being cooled at which the stream will be liquefied and the specific heat will be relatively constant over the temperature range existing in the exchanger. An expander inlet temperature is then normally chosen which will result in an expander outlet temperature near the dew point of the expanded gas. The fraction of the total gas flow sent through the expander which will result in minimum temperature differences in the heat exchangers is then determined.

The design of an efficient expander cycle requires elimination of irreversibilities in the process by maintaining small temperature differences in the heat exchanger, high efficiencies in the expansion, resulting in low entropy changes in the gas passing through the expanders, and efficient utilization of the work produced by the expansion. Some of the process variables which affect the cycle efficiency are operating pressures, expansion stages, expansion efficiency of the turbine or expansion engine, and subcooling of liquid before expansion.

An advantage of the expander cycle over the cascade cycle is simplicity, which can result in lower capital costs for very small plants, at the expense of fuel costs. For plants of the size normally anticipated for peak-shaving of 10^6 CF/day and up, the added energy requirements result in higher compressor costs, which, along with the expander investment and added fuel costs, more than negate the savings in heat exchangers, controls, and refrigerant systems.

However, the expansion engine cycle has definite application in situations where considerable amounts of gas are being let down in pressure through a valve into distribution mains or power stations. In these situations, it is possible to utilize this energy by reducing the gas pressure through a turbine, and removing the refrigeration produced to liquefy gas. As shown in Figure 2.1.2.-10, the expanded vapors, after being warmed against high-pressure gas, can be sent to the low-pressure distribution system or power station, thus reducing or eliminating recycle compression. This arrangement can result in savings in investment and operating costs; however, the flow and pressure drop must be available for a large percentage of the year to make the system economical.

The expander cycle was applied successfully for producing LNG on the Constock barge at a design capacity of 7 million CF/day. The expander cycle was selected primarily because of the need for a light and compact refrigeration unit which could be mounted on a barge. The barge cycle, shown in Figure 2.1.2.-13, was adaptable to the use of lighter and more compact centrifugal machinery, and had fewer components than a cascade cycle. It was recognized that the cycle was less efficient and resulted in greater installed horsepower than a cascade system. The Freon and recycle compressors were driven with steam turbines, and the fuel consumption was about 30% of the total feed gas.

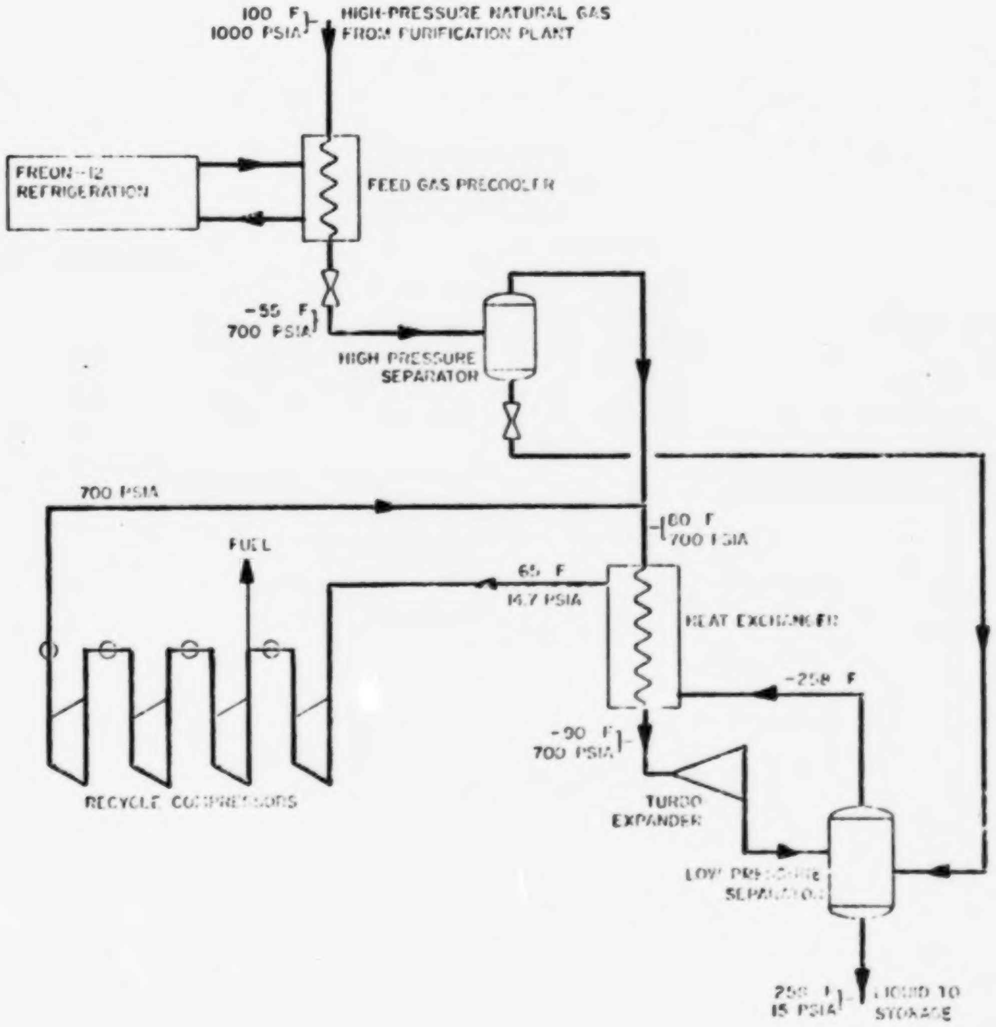

100 F
1000 PSIA
HIGH-PRESSURE NATURAL GAS FROM PURIFICATION PLANT
FREON-12 REFRIGERATION
FEED GAS PRECOOLER
-55 F
700 PSIA
HIGH PRESSURE SEPARATOR
700 PSIA
FUEL
80 F
700 PSIA
65 F
14.7 PSIA
HEAT EXCHANGER
-258 F
-90 F
700 PSIA
RECYCLE COMPRESSORS
TURBO EXPANDER
LOW PRESSURE SEPARATOR
258 F
15 PSIA
LIQUID TO STORAGE

Figure 2.1.2.-13. EXPANDER LIQUEFACTION CYCLE
(Capacity: 7 Million CF/Day at 7000 bhp)[43]

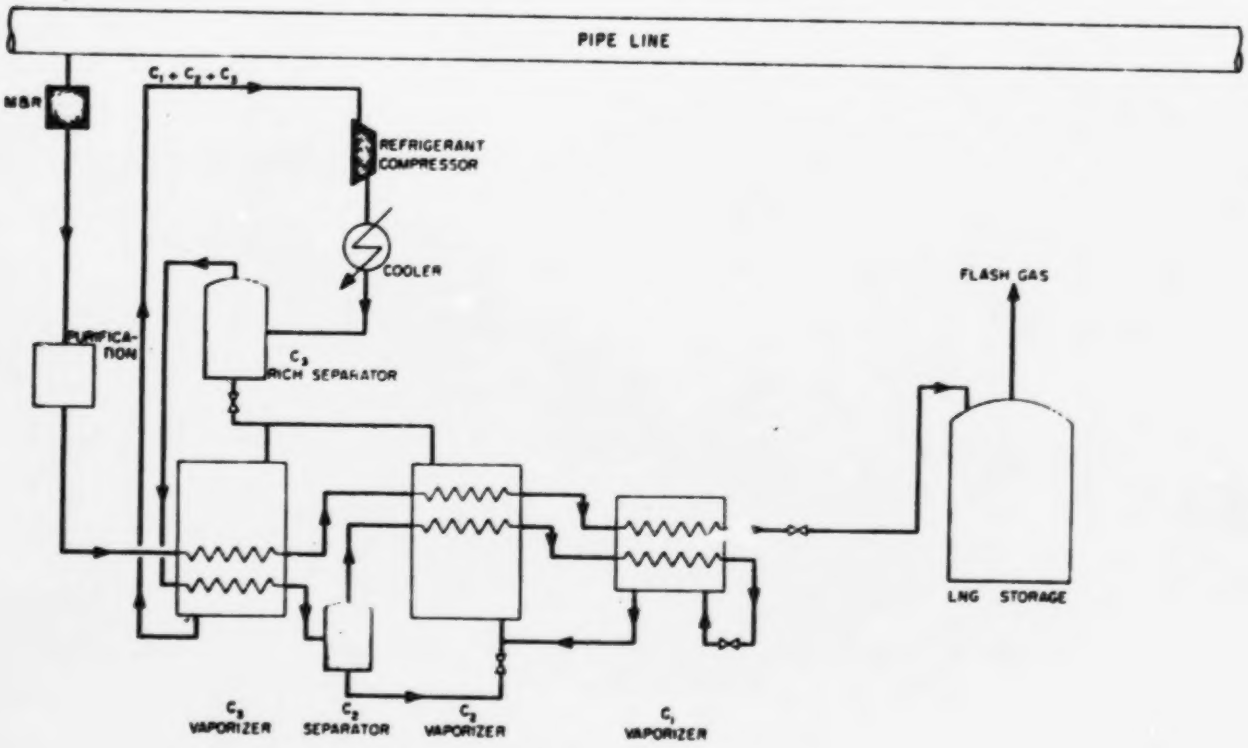

PIPE LINE
MBR
$C_1 + C_2 + C_3$
REFRIGERANT COMPRESSOR
COOLER
PURIFICATION
C_3 RICH SEPARATOR
FLASH GAS
LNG STORAGE
C_3 VAPORIZER
C_2 SEPARATOR
C_2 VAPORIZER
C_1 VAPORIZER

Figure 2.1.2.-14. INCORPORATED CASCADE CYCLE FOR NATURAL GAS LIQUEFACTION

Axial compressors that have efficiencies of 0.88 can be built up to 80,000 hp. Gas turbines in the 20,000-hp range have been used for continuous service applications. Four of them could be paralleled to provide power for a single compressor. The waste-heat recovery equipment is optional because the natural gas can be available at 10¢/million Btu and, at this low energy cost, relatively low-cycle thermal efficiency is not critical. Both methods and their efficiencies are illustrated in Figures 2.1.2.-15 and 2.1.2.-16. Figure 2.1.2.-17 illustrates a method of compounding gas turbines to achieve 80,000 hp in direct gas turbine drive to a 2-stage natural gas compressor having a 200-psia suction and a 1000-psia discharge. The exhaust gas fires a boiler. The steam from both boilers is fed into a 60,000-hp steam turbine. The expansion turbine is mechanically coupled to this line shaft, which powers the same size compressors used in the gas turbine branch. The additional four gas turbines are not shown, but their exhaust line is indicated. This complete branch is capable of generating 240,000 hp because it consists of eight gas turbines, one steam turbine, and one expander. Before analyzing some of the major costs of this installation, we will consider the expander liquefaction plant in relation to a single 10,000-MW power plant facility. A heat rate of 10,000 Btu/kWhr may not be too conservative when accounting for pumping and transportation losses. Using this assumption, 2400 million CF/day of one million Btu/CF vaporized natural gas would be required to power a 10,000-MW power generating block. Thus, exactly 10 plants of the size indicated in Figure 2.1.2.-18 would be required.

It is our estimate that the plant construction will be approximately $200 per installed horsepower. This includes the steam turbine branch. Thus, a 240,000-hp 240 million CF/day plant would cost about $48 million.

FUEL HEAT VALUE, %	100
GENERATOR, %	23.5
STACK LOSS, %	74.5
MISCELLANEOUS LOSSES, %	2.0
TURBINE TEMPERATURE, °F	1500
STACK TEMPERATURE.	750
ENGINE THERMAL EFFICIENCY, %HHV	23.5
ELECTRICAL HEAT RATE, Btu/hp–hr	10,800

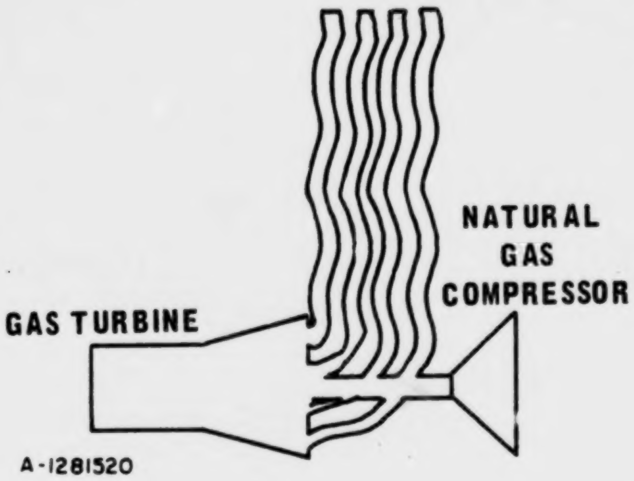

NATURAL
GAS
COMPRESSOR
GAS TURBINE
A-1281520

Figure 2.1.2.-15. BASIC POWER UNIT [58]

FUEL HEAT VALUE, %	100
STACK LOSS, %	37.3
WASTE–HEAT COMPRESSOR, %	9.0
MAIN COMPRESSOR, %	23.5
CONDENSER LOSS, %	27.2
MISCELLANEOUS LOSSES,	3.0
TURBINE TEMPERATURE, °F	1500
STACK TEMPERATURE, °F	370
PLANT THERMAL EFFICIENCY, %HHV	32.5
PLANT HEAT RATE, Btu/hp–hr	7,850

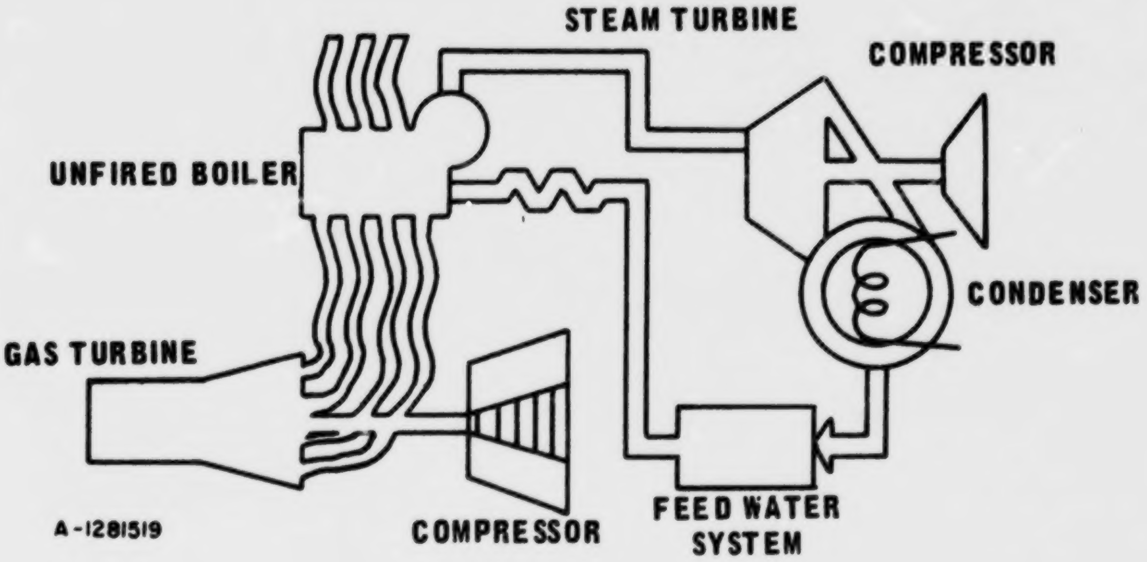

STEAM TURBINE
COMPRESSOR
UNFIRED BOILER
CONDENSER
GAS TURBINE
A-1281519
COMPRESSOR
FEED WATER SYSTEM

Figure 2.1.2.-16. BASIC POWER UNIT WITH HEAT RECOVERY[58]

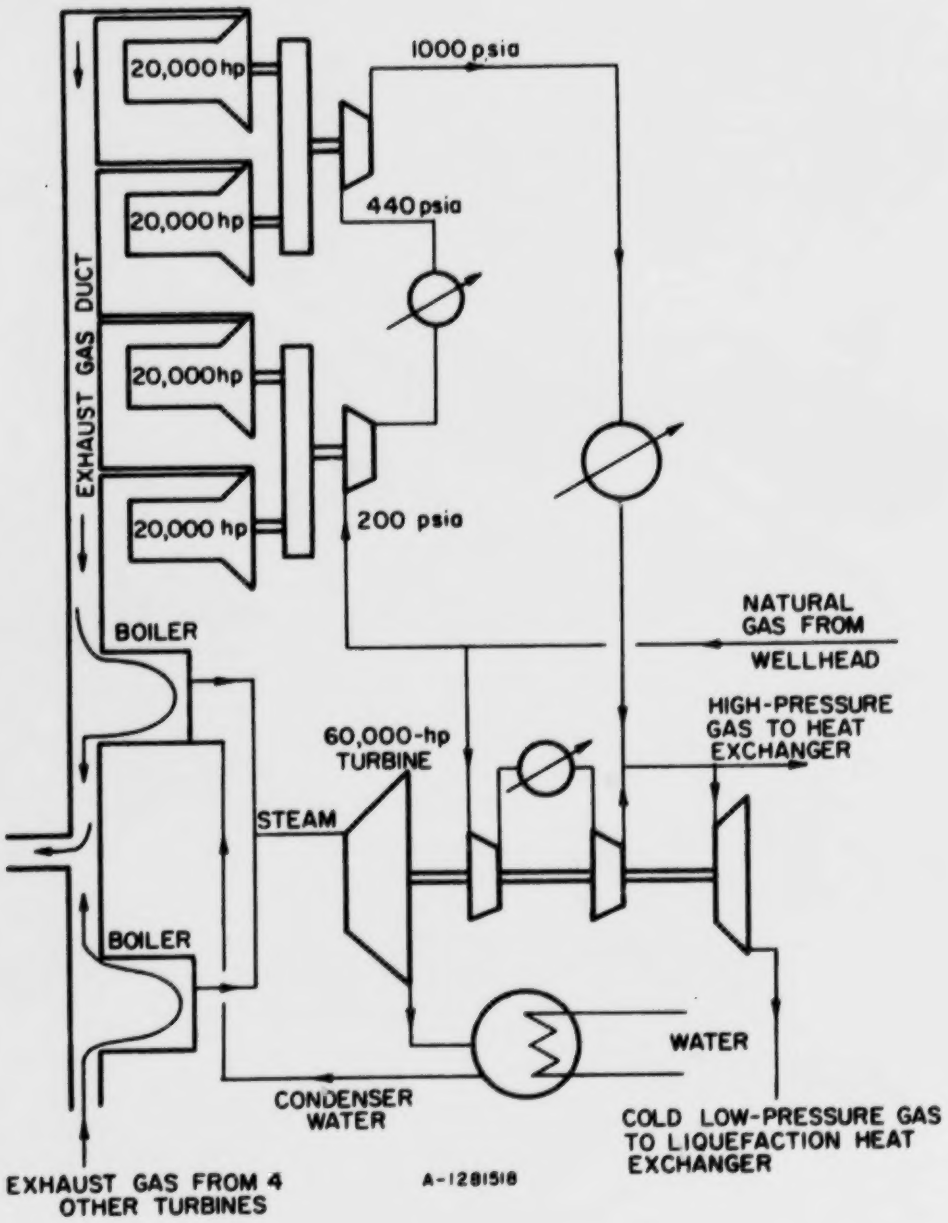

20,000 hp
20,000 hp
20,000 hp
20,000 hp
EXHAUST GAS DUCT
1000 psia
440 psia
200 psia
NATURAL GAS FROM WELLHEAD
BOILER
HIGH-PRESSURE GAS TO HEAT EXCHANGER
60,000-hp TURBINE
STEAM
BOILER
WATER
CONDENSER WATER
COLD LOW-PRESSURE GAS TO LIQUEFACTION HEAT EXCHANGER
EXHAUST GAS FROM 4 OTHER TURBINES
A-1281518

Figure 2.1.2.-17. 240,000-hp POWER TRAIN FOR EXPANSION CYCLE

The expander cycle has been applied to the design of an LNG peakshaving facility for the San Diego Gas & Electric Company. In this case, the cycle utilizes the energy available in the existing pipeline supply of gas to a power plant, where the pressure is being reduced from approximately 300 psia to approximately 80 psia. Approximately 25 million CF/day of feed gas enters the plant, 2 million CF/day of which will remain in storage as liquid at atmospheric pressure, while 23 million CF/day will flow to the power plant at approximately 80 psia.

2. 1. 2. 3. 2. Investment Costs

The expansion cycle as illustrated in Figure 2. 1. 2. -10 does not attempt to recover any energy from the expansion engine. Such a cycle is justified only where small peakshaving plants are being considered. It would not be considered practical for large base-load plants. However, unlike Figure 2. 1. 2. -13, which also does not recover the expansion engine power, the former method splits off part of the main stream for direct Joule-Thomson expansion. Point 3 of Figure 2. 1. 2. -11 is in the liquid zone where the Joule-Thomson expansion is accomplished with a minimum of irreversibilities.

The logical result of these two approaches is illustrated by Figure 2. 1. 2. -14. However, all of these three variations on the expansion engine require approximately twice the power that the cascade cycle does. This means approximately 1000 hp/million SCF/day, assuming that pipeline pressure into the liquefaction plant is at least 300 psia. Based on a base-load plant of 400 million CF/day, this is 400, 000 hp.

2. 1. 2. 3. 3. Operating Costs

The 240, 000-hp expansion cycle rated at 240 million CF/day based on 300 days/yr has an annual capacity of 72 billion CF. Assumptions will be the same as used for Figure 2.1.2.-8 except, of course, the fuel consumption will be based on 1000 hp required for 1 million CF/day and a heat rate of 7850 Btu/hp-hr as shown in Figures 2. 1. 2. -15 and 2. 1. 2. -16.

This cost of 20.2¢/1000 CF (Table 2.1.2.-6 is the lowest cost of the systems surveyed thus far. The estimate of the capital cost was not made from vendors' bids. Since this is such an important factor, it should be noted that this amount could be higher or lower. The scale effect is not deemed important here in comparing the results with those of Figure 2. 1.2. -8 which are for a smaller plant.

Table 2. 1. 2. -6. CALCULATION OF UNIT COST FOR LIQUEFYING 72 BILLION CF OF NG BY EXPANSION CYCLE

1. Liquefaction Facility Investment	$48, 000, 000
2. Capital Cost	
Annual Principal	$ 3, 200, 000
6% Average Interest	$ 1, 444, 000
6% Tax Rate and Insurance	$ 2, 880, 000
Total	$ 7, 524, 000
3. Sale Gas	
(5. 45¢/1000 CF) X (72 X 10^9 CF)	$ 3, 920, 000
4. Fuel Gas	
(240, 000 hp) X (300 days/yr) X (24 hr/day) X (7850 Btu/hp-hr) X (05. 45¢/1000 CF) X (1000 CF/ 1. 1 X 10^6 Btu)	$ 672, 000
5. Annual Operating Expense	
5% of $48, 000, 000	$ 2, 400, 000
6. Total Annual Cost	$14, 516, 000
7. Unit Cost of LNG	20.2¢/1000 CF

2.1.2.4. Combined Cycles

2.1.2.4.1. Process Description

Inherent in the cascade and expansion cycles is the Joule-Thomson effect. The Joule-Thomson effect occurs in the last stage in each cycle and describes the cooling effect on a liquid when it is brought from a region of high pressure to a region of low pressure. It is natural then to expect certain advantages in combining all three refrigeration techniques. Figure 2.1.2.-18 illustrates the combined cycle.

A 200-psia feed stream (A) joins a recycle stream (B) at that pressure. The combined stream (D), shown as 100 lb of gas, is compressed in a 2-stage compressor (C-1) to 1000 psia. Water cooling (not shown) is used to maintain gas stream temperature at not greater than 105°F. This stream (F) is fed into a heat exchanger (E-1), then into exchanger (E-2) where propane cooling from a separate mechanical refrigerator (not shown) brings the gas stream (H) temperature to –15°F. Further cooling in the heat exchanger by the recycle gas cools the 1000-psia stream (G) to –70°F, at which time 86.7% of the stream is extracted and passes through an expander (J). The expander discharges a mixture (K) of 14.5% (by weight) liquid and 74.1 lb (the balance) of vapor into a separator (S-1). Vapor stream (M) at –178°F recirculates back through the heat exchanger (E-1) eventually returning to compressor stream B at + 95°F and 200 psia. The balance of the high-pressure (1000-psia) stream (N) which now leaves the heat exchanger at –178°F is expanded through a Joule-Thomson valve (V-1) down to atmospheric pressure. This cools the stream (P) to –258°F and liquefies approximately 68% of it. The same percentage of liquid at atmospheric pressure is formed at R from the 12.6 lb of liquid (L) leaving the separator (S-1) as it expands through the Joule-Thompson valve (V-2) from 200 to 14.7 psia.

The cold gas at –258°F is returned via stream T back to the heat exchanger (E-1) while 17.5 lb of LNG is removed from separator S-2 through line W. A separate recycle compressor (C-2) boosts stream Y

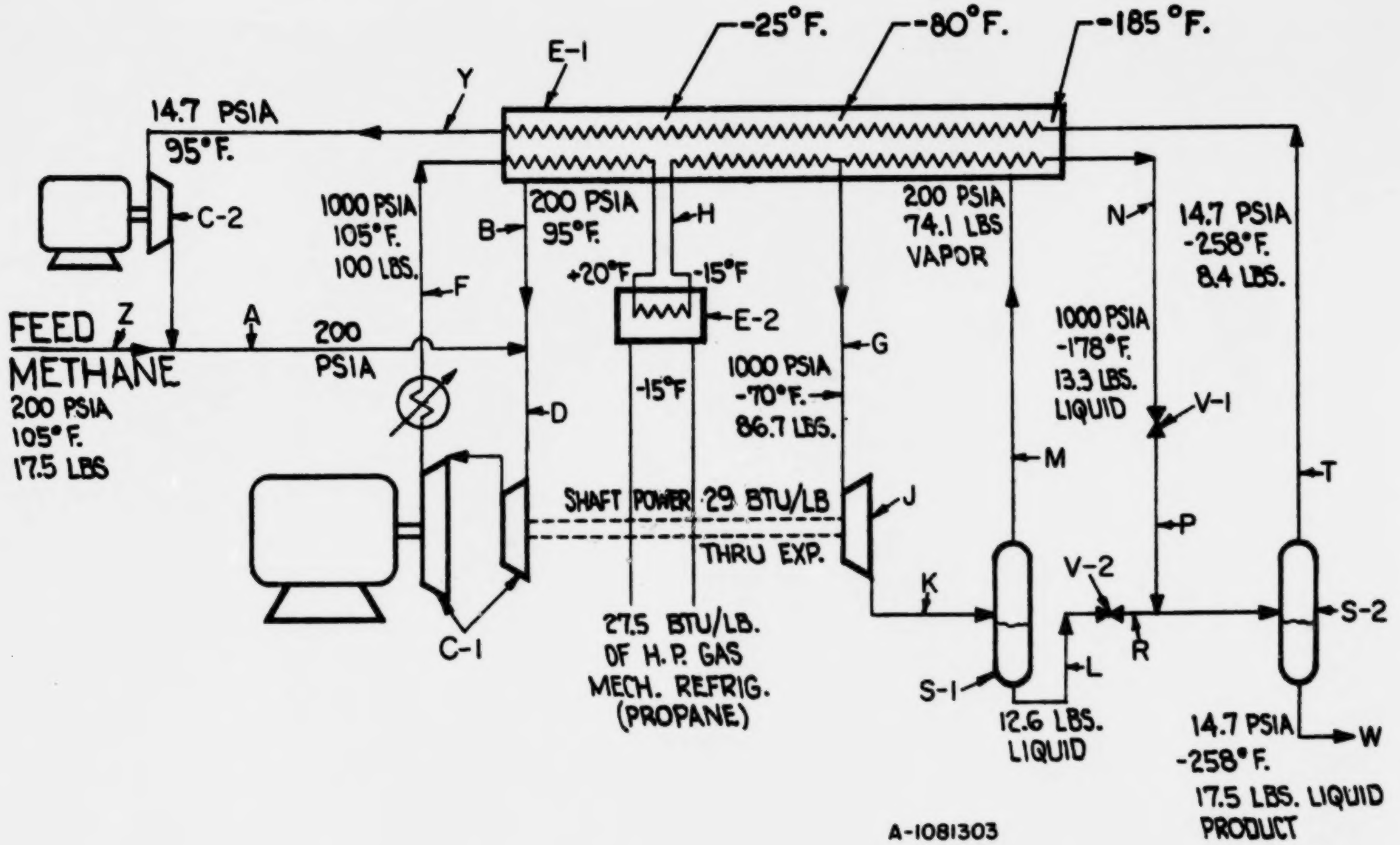

E-1
-25°F.
-80°F.
-185 °F.
Y
14.7 PSIA
95°F.
C-2
1000 PSIA
105°F.
100 LBS.
B
200 PSIA
95°F.
H
+20°F
-15°F
E-2
-15°F
F
FEED
Z
A
200
PSIA
METHANE
200 PSIA
105°F.
17.5 LBS
D
C-1
SHAFT POWER 29 BTU/LB
THRU EXP.
27.5 BTU/LB.
OF H.P. GAS
MECH. REFRIG.
(PROPANE)
G
1000 PSIA
-70°F.
86.7 LBS.
J
K
200 PSIA
74.1 LBS
VAPOR
M
S-1
L
12.6 LBS.
LIQUID
N
1000 PSIA
-178°F.
13.3 LBS.
LIQUID
V-1
P
V-2
R
14.7 PSIA
-258°F.
8.4 LBS.
T
S-2
14.7 PSIA
-258°F.
17.5 LBS. LIQUID
PRODUCT
W
A-1081303

Figure 2.1.2.-18. FLOW DIAGRAM FOR METHANE LIQUEFACTION BY USE OF COMPRESSION, TURBOEXPANSION, AND MECHANICAL REFRIGERATION [67]

back up to 200 psia. It then joins the 17.5-lb makeup stream (Z) from the pipeline.

The expander (J) is mechanically coupled to the primary compressor (C-1). Its work contribution to the cycle is not a major factor in the overall power consumption.

These processes are illustrated graphically in Figure 2.1.2-19. The enthalpy-pressure diagram shows the rather large blocks of heat that must be removed at high temperatures, but does not indicate the importance of the low-temperature portion of the cycle and its effect on the work input.

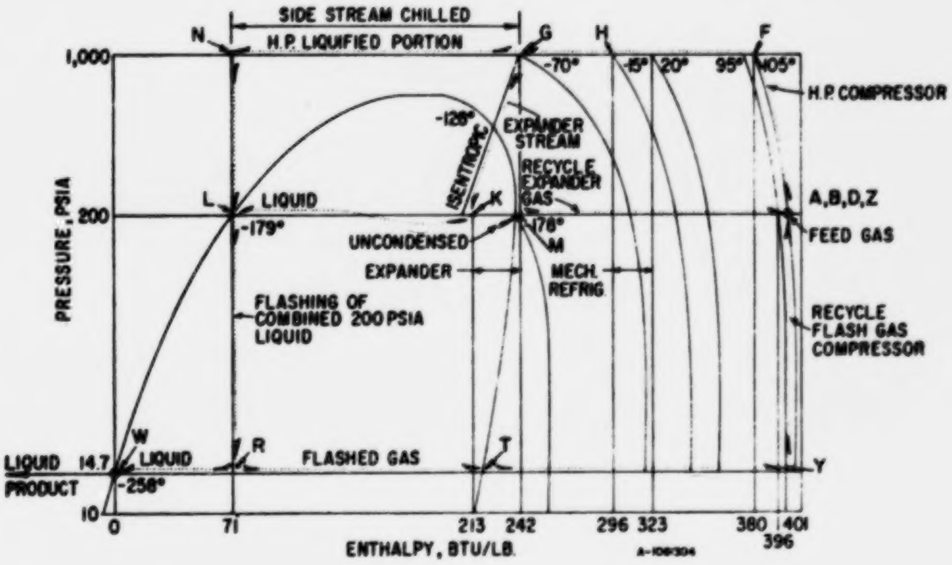


Figure 2. 1. 2. -19. GRAPHICAL CYCLE DESCRIPTION [66]

The isentropic expansion is theoretical since expander efficiency is actually approximately 85 %. Various refinements can be included in this cycle, including optimizing each temperature and pressure for minimum work input.

This method of NG liquefaction has not been used on any large LNG base-load liquefaction system. Its efficiency is not as high as that of the cascade cycle because a large part of the refrigeration is achieved through the Joule-Thompson effect. This combined cycle does offer some simplification over the three-component cascade (Figure 2.1.2.-6

2. 1. 4. 2. Investment Costs

The combined cycle shown in Figures 2. 1. 2. -18 and 2. 1. 2. -19 is capable of producing 1 million CF/day of LNG when absorbing 13, 200 hp-hr of energy. The primary reason for this improvement over the simple cascade is the more extensive use of heat exchangers; however, these exchangers add significantly to the cost of installation. There is a point at which the capital cost of the heat exchanger and its effect on improved efficiency results in an optimum size of compressor and hence a minimum unit cost of LNG. To find this point would require a more detailed design study and would include actual quotations. It would also require a firm price for the cost of gas rather than an assumed one, such as the value of 6¢/million Btu that we have used throughout our study.

Because of the close similarity between the combined cycle and the expansion cycle, it is of value to examine plants of the same sizes. The estimated cost of the combined cycle system for a plant liquefying 72 billion CF/yr is $55 million. The cost of "fuel gas" will be taken as 15% of the cost of "sale gas."

2. 1. 2. 4. 3. Operating Costs

Table 2. 1. -7 outlines the cost factors for this type of plant. A comparison of the combined cycle and the expansion cycle indicates that the expansion cycle is 10% less expensive than the combined cycle. The importance of the capital items is once again shown. It should be emphasized that the closeness of these figures requires that bids be taken if a true comparison is to be made.

Table 2.1.2.-7. UNIT COST OF LIQUEFYING 72 BILLION CF/YR OF NG BY COMBINED CYCLE

1. Liquefaction Facility Investment	$55,000,000
2. Capital Cost	
Annual Principal	$ 3,667,000
6% Average Interest	$ 1,650,000
6% Tax Rate and Insurance	$ 3,300,000
Total	$ 8,617,000
3. Sales Gas	
(5.45¢/1000 CF) X (72 X 10^9 CF)	$ 3,920,000
4. Fuel Gas	
15% ($3,920,000)	$ 588,000
5. Annual Operating Expense	
5% of $55,000,000	$ 2,750,000
6. Total Annual Cost	$15,875,000
7. Unit Cost	22.1¢/1000 CF

2.1.2.5. Miscellaneous Cycles

2.1.2.5.1. Process Description

2.1.2.5.1.1. Liquid Nitrogen Exchange

The liquefaction of natural gas or methane can be accomplished by simple heat exchange with a medium at a lower temperature than −258°F. Nitrogen at −320°F and oxygen at −297°F are two common liquids whose temperatures at atmospheric pressure are below that of methane. Figure 2.1.2.-20 illustrates a schematic arrangement of such a liquefaction method.

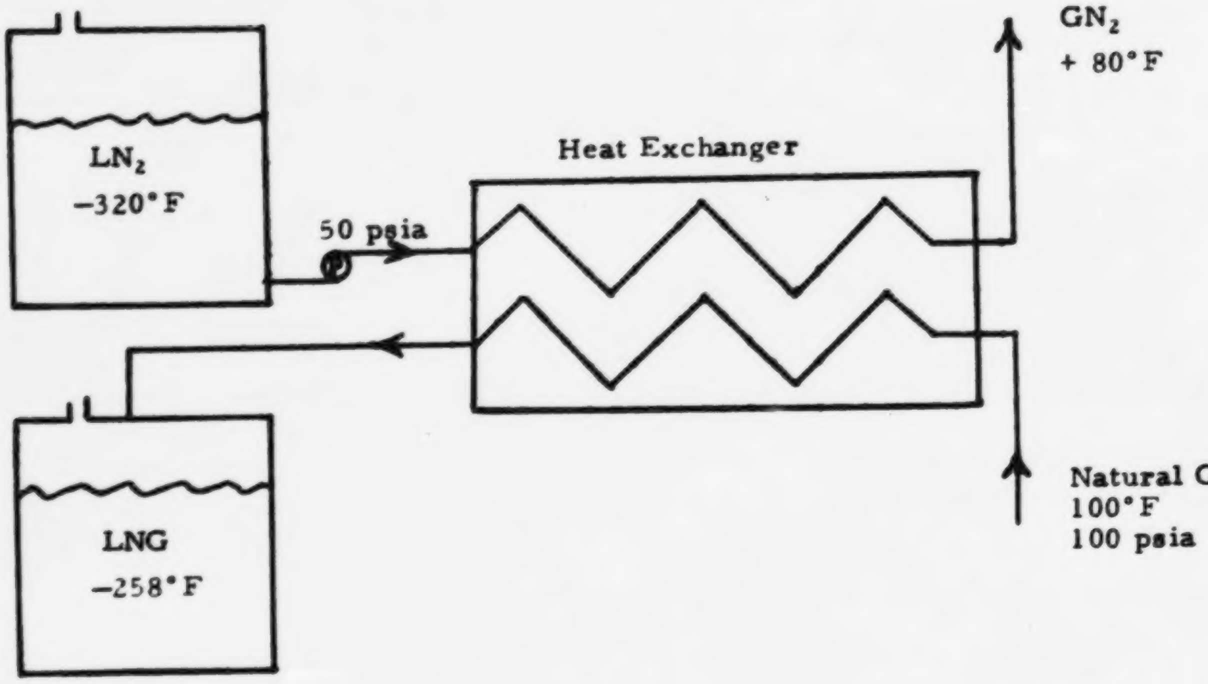


Figure 2.1.2.-20. SCHEMATIC DIAGRAM OF LIQUID NITROGEN EXCHANGE

The theoretical cooling required to convert 1 lb of natural gas (assumed as 100% methane for simplicity) at 100°F and 100 psia to a liquid at atmospheric pressure at −258°F is 398 Btu. Liquid nitrogen at −320°F and 50 psia may be discharged through a countercurrent heat exchanger. Assuming LN_2

exhausts at +80°F, it will extract 186.5 Btu/lb LN_2 from the LNG. Thus, 2.14 lb of LN_2 are required to liquefy 1 lb of LNG.

This is a very inefficient process from the standpoint of the large temperature differential between the two cold ends of the exchanger. If desired, subcooling of the LNG can be achieved. The solidification point of LNG is approximately −296°F. A refinement of this method would be to precool the natural gas and thus reduce the amount of LN_2 required. The cooling effect of the LN_2 is reduced, but, because the specific heat of methane gas (0.52 Btu/lb-°F) is much higher than that of nitrogen (0.24 Btu/lb-°F), less liquid nitrogen is used.

This method is used for small-scale laboratory purposes, but has been proposed for large base-load vaporization as well. In the latter case, its proponents state it could be used as part of a base-load vaporization system

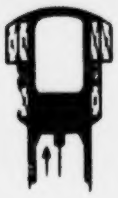
2.1.2.5.1.2. Cold Gas Refrigerator (CGR)

This type of liquefaction technique is presently limited to smaller sizes and is available commercially in practical sizes up to 2.5 million CF/day. The commercial units are manufactured in The Netherlands and marketed in the U.S. by Norelco and CVI Corp.

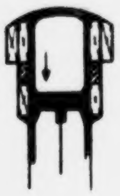
The CGR operates on a closed-system gas refrigeration cycle. The cylinder houses a compression and a displacer piston — both coupled to a single two-throw crankshaft by separate connecting rods. The crankshaft produces sinusoidal-phased harmonic motions of the two pistons, which alternately compress, transfer, and expand the working fluid within the cylinder.

The displacer piston divides the space above the compression piston into two sections: The space just beneath the displacer piston is used for compression while the space above the displacer piston is used for expanding and cooling the working fluid. These sections are connected by a channel containing three heat exchangers: water cooler, regenerator, and condenser. The working medium within the cylinder is transferred from the compression space to the expansion chamber and back again by the motion of the displacer.

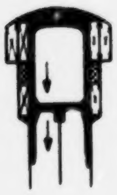
The process of compression and expansion during one revolution of the crankshaft may be described by the following four phases.



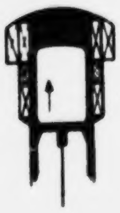
Phase 1. The working medium is contained mainly in the compression space where it is compressed by the main piston to maximum pressure.



Phase 2. The displacer moves the compressed gas from the compression space to the expansion chamber, the gas flowing through the cooler, and the regnerator to the condenser head. The heat of compression is carried off by the water when the gas passes through the cooler. The gas is further cooled in its passage through the regenerator.



Phase 3. In this phase the temperature of the gas contained in the expansion space is reduced by the expansion effected by the simultaneous downward movement of the piston and the displacer. The low temperature produced in the condenser head is transferred to the condenser product.



Phase 4. The cold gas is driven back to the compression space by the return motion of the displacer. In passing back to the compression space, the working medium absorbs the heat stored in the regenerator during the second phase.

The basic principles used in the CGR are essentially the same as in the combined cycle except that the compression, expansion, and heat exchange occur in the same cylinder head. The other difference is that a sealed working fluid – either hydrogen or helium gas – is used as the cooling medium. It is compressed and expanded, always remaining in the gaseous form, until it cools the head of the cylinder to a suitable liquefaction temperature. The natural gas is piped into the final heat exchange area (the head of the cylinder) at about 100 psig where it liquefies on contact with the cold head.

A single-cylinder skid-mounted CGR is shown in Figure 2.1.2.-21. The complete CGR LNG cycle is shown in Figure 2.1.2.-22.

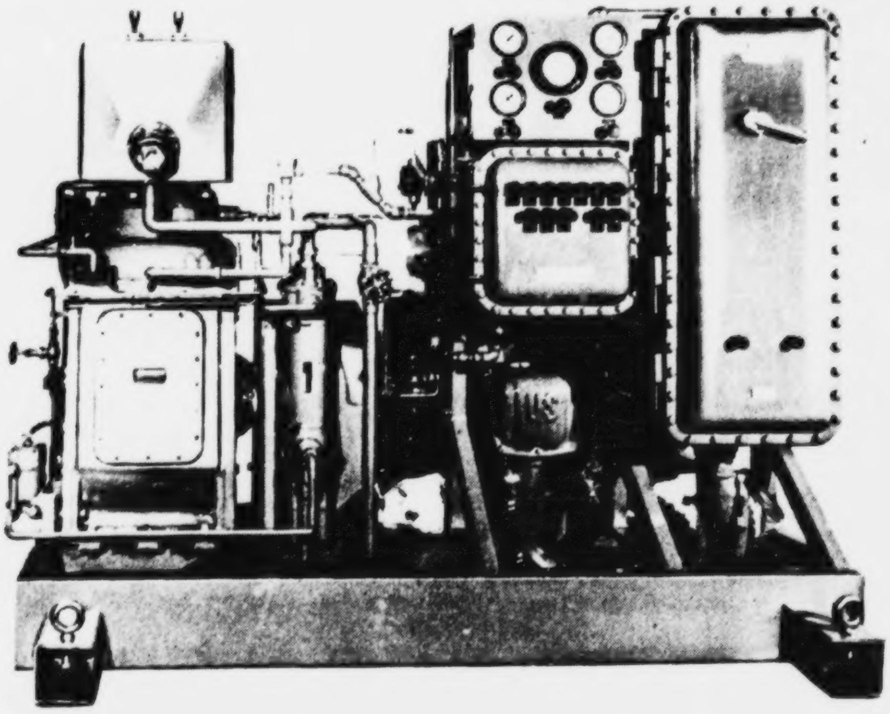


Figure 2.1.2.-21. SKID-MOUNTED CGR WITH DRIVE[72]

The important components are –

1. Precool

   A Freon refrigerator operating in the –60° to –70°F range provides an efficient supply of refrigeration for precooling the CGR inlet stream.

2. Heat Exchanger

   A counterflow-type heat exchanger would be used to recover the refrigeration effect available in the storage tank return stream. By heat exchange with the warm supply gas, the return stream temperature would be warmed to ambient, thereby providing effective precooling of the inlet stream.

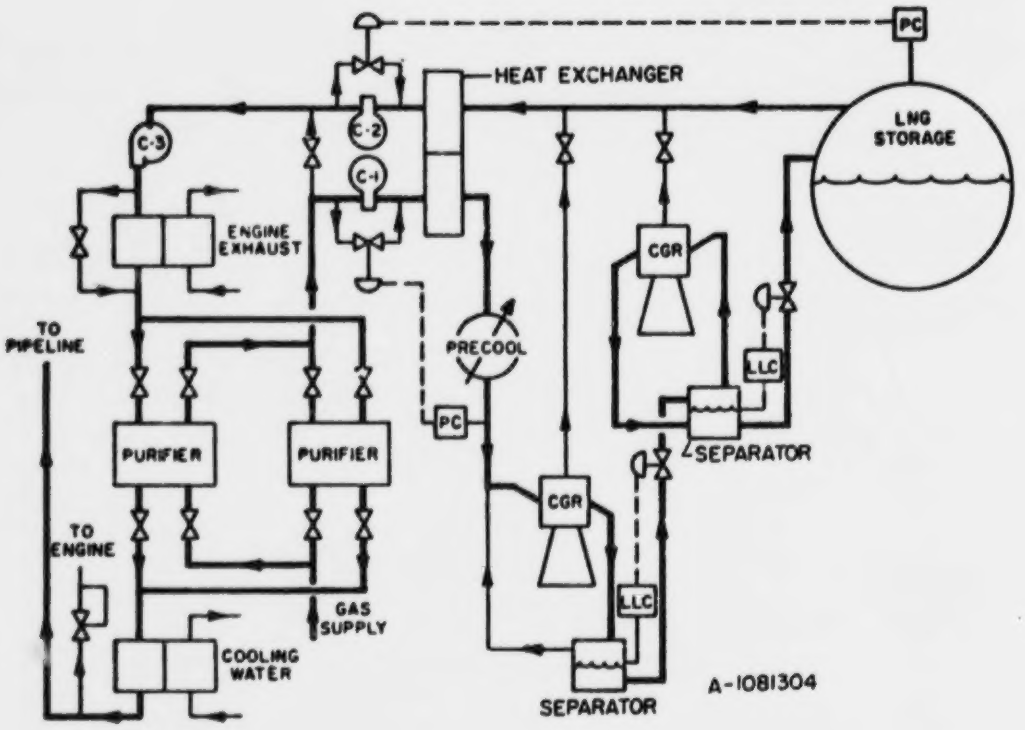

PC
HEAT EXCHANGER
LNG STORAGE
C-3
C-2
C-1
ENGINE EXHAUST
TO PIPELINE
PRECOOL
CGR
LLC
SEPARATOR
PURIFIER
PURIFIER
PC
TO ENGINE
GAS SUPPLY
COOLING WATER
CGR
LLC
SEPARATOR
A-1081304

Figure 2. 1. 2. -22. COMPLETE LNG CYCLE WITH CGR[72]

3. Supply Stream Compressor, C-1

For larger capacity systems requiring multiple-cylinder installations, it is desirable from an operating efficiency standpoint to operate the cylinders in a series arrangement condensing at two different temperature levels with intermediate Joule-Thompson expansion. A supply pressure of 400 to 500 psig is desirable to take full advantage of this technique. If line pressures are relatively low, a booster compressor (C-1) would be installed in the supply stream.

4. Compressors C-2 and C-3

The nitrogen-enriched storage tank return stream is used for purifier regeneration and as fuel in the gas engine driver. A compressor (C-2) is installed to boost the pressure of this return stream from storage tank pressure. The required adsorber regeneration flow rates are provided by mixing the storage tank return stream with cleaned gas from the supply. An increase in the gas supply pressure level is supplied by C-3. (A common frame would normally be used for both Compressors C-1 and C-2.)

5. Separator Tanks

Operation with a multiple-cylinder arrangement in series requires separator tanks with suitable instrumentation and controls.

6. Purifier

Purifiers using a molecular sieve with provisions for suitable regeneration are required. A gas engine exhaust provides the heat source for the purifier.

7. Transfer System

When a CGR is installed to operate in conjunction with an existing plant, the gas to be condensed is often at a low pressure. If an elevation difference is not available between the CGR unit and the storage tank liquid, a transfer system is required.

2.1.2.5.2. Investment Costs

2.1.2.5.2.1. Liquid Nitrogen Exchange

By making some changes in the process outlined in Figure 2.1.2.-3 this method can be much more effective. The liquid nitrogen (LN_2) pump pressure is increased to 1000 psia instead of 50 psia. The LN_2 is pumped into the heat exchanger where it is warmed by natural gas to 80°F. This warmed nitrogen gas is now expanded in a turbine. During expansion it is cooled to −230°F and 25 psia (assuming 85% expander efficiency). This cool nitrogen gas is returned to the heat exchanger where it extracts an additional 75 Btu/lb of nitrogen from the incoming natural gas stream. The pump work adds only 5 Btu/lb of nitrogen. This cooling effect is additive to that of the original vaporization of LN_2, which is now 180 Btu/lb instead of 186.5 Btu/lb. As a result of this increased cooling ability, only 1.6 lb of LN_2 is theoretically required to liquefy 1 lb of LNG. This can be compared to the 2.14 lb of LN_2 required by the nonexpansion method.

The expander also produces 69 Btu/lb of LN_2 of useful work. For a 200 million CF/day NG liquefaction plant, this power is equivalent to 16,300 hp, which amply covers the 1080 hp required for pumping, but has only a minor effect on the overall liquefaction plant economics. The major part of the capital equipment for such a plant is at the nitrogen production facility. This will be discussed in more detail in Section 2. 2. 2. on technological trends of liquefaction cycles because this system would be new and has not been clearly defined.

The major NG liquefaction components required are the LN_2 pumps, the heat exchanger, the expansion turbine, and the LNG tanks. By a judicious use of compartments and shore storage tanks, it is possible to use only one additional storage tank to store the LN_2 and LNG products. There may be some dilution difficulties in immediately filling a tank with LN_2 if the tank has just been emptied of LNG, and vice versa, but for the sake of this study we shall neglect them. Based on the same production, storage capacity for the cryogenic shore storage would be increased by one tank at each site. Therefore, to obtain a truer picture of the cost of this process, two additional 300, 000-bbl tanks will be included as part of the capital investment. The LNG tanks are not included in the other analyses, and this practice is retained for this analysis as well.

Another large capital investment is in the LN_2 plant that is required at the LNG vaporization site. Typically, the LNG would be transported to the vaporization site where some cold recovery technique would be employed to cool the air, liquefy it, and separate the LN_2. A detailed economic analysis of cold recovery techniques is outside the scope of this study. However, the "cold recovery" investment factor can be estimated and will be included in the operating costs section, where the LN_2 will be treated as a "fuel."

The LN_2 produced at the LNG vaporization site is then transported in the previously emptied LNG tankers to the gas pipeline where LNG liquefaction occurs. Assuming a liquefaction plant site is in Venezuela, approximately 2000 miles from New York and Philadelphia ports, establishes the number of storage tanks required for the 200 million CF /day liquefaction plant. We estimate the installed cost for this facility at $16 million.

This includes 300,000 bbl of storage at $10/bbl at both the gas field site and the LN_2 liquefaction site. This is a total of 600,000 bbl of storage. This cost also includes heat exchangers, turbine and electric generators, pumps, piping, and land improvements. No pipelines, piers, or other storage tanks are included.

In view of this very low investment cost, it should be emphasized that the greater part of the investment is at the LNG vaporization site and that this system is more capital intensive than it appears at first glance.

2.1.2.5.2.2. Cold Gas Refrigeration

If we consider Figure 2.1.2-22 as typical — a pair of 4-cylinder cold gas refrigerators — this system will require a 600-hp electric motor and be capable of liquefying 0.6 million CF/day. The 4-cylinder CGR is the largest size CGR made; two of them could be mounted together on a skid. Four such units would be capable of liquefying 2.5 million CF/day and would appear to be the practical limit for the CGR system. These units would also require compressed natural gas at a pressure of between 500 and 600 psia as well as a Freon refrigeration precool that would lower the temperature of the gas going into the first-stage CGR to −60°F. The installed cost overseas for such a system less tankage, using electric motor drive and with all other accessories, is approximately $2 million.

Although present-size units and general configuration of the reciprocating components indicate that the CGR will never meet the liquefying needs of the large base-load system under consideration, it would be well to examine the cost parameters as part of a shipboard boil-off reliquefaction system or a tank boil-off reliquefier on shore.

2.1.2.5.3. Operating Costs

2.1.2.5.3.1. Liquid Nitrogen Exchange

As has been shown under Investment Costs, this system requires 1.6 lb of LN_2 to produce 1 lb of LNG. Allowing for transportation and other heat losses during the 3-1/2-day sea journey, the tanker would load 1.7 lb of LN_2. It could load up to 1.9 lb of LN_2 /lb LNG returned because this is the ratio of the specific gravity of the two liquids.

Cold recovery techniques would be used to produce LN_2. We estimate that the cost of LN_2 under these conditions will be approximately 0.25¢/lb. This is not an easy number to estimate because the actual technique of cold recovery has not been finalized. Our figures indicate that relatively little energy is needed to compress air for the air liquefaction and separation cycle because virtually all the refrigeration will be derived from the vaporized LNG. The distillation column and cold box may be different from those employed by conventional plants. Also, there are unprecedented factors of scale, even for this modest size 200 million CF/day plant. The liquid nitrogen facility is equivalent to 7100 tons/day of production, which would be the largest LN_2 plant in the world. Scale factors are very important here, as are duty-cycle factors. Most commercial plants must be designed with some flexibility of output ratio between the oxygen and nitrogen and between the ratio of gases and liquids. A relatively narrow range would appear to be suitable here.

Approaching this cost 0.25¢/lb of LN_2 from the other standpoint we see from Table 2.1.2.-8 that it amounts to a cost of $12,950,000. Almost all of this cost is in the form of amortization expense, and thus represents a very sizable capital investment. (Perhaps our estimate here is on the high side.) In any event, this cost is not accurately definable at this time and, unfortunately, it is the key to the economics of this liquefaction system.

Table 2.1.2.-8 gives the derivation of the unit cost of gas using this method. The cost of 28.6¢/1000 CF is one of the highest for the systems analyzed. A 7% operating expense figure was used instead of the 5% used in the other systems because of the greatly reduced capital expense in this case. Another aspect of this plant is the availability of 15,120 hp of excess energy from the nitrogen turbine. A net credit of 5 mills/kWhr is taken on the assumption that the turbine will drive an electric generator. The generator rating is 10,000 kW.

In concluding this analysis it is well to emphasize the high degree of uncertainty of the cost of nitrogen. The LN_2 exchange method does have the advantage of a minimum capital investment on a foreign shore, which can be a great advantage. Credits for oxygen and nitrogen gas values are also neglected because of the likelihood that large markets for these by-products are not available. However, there is a probability that some credits can be derived from both of these by-products.

Table 2.1.2.-8. CALCULATION OF UNIT COST OF LNG
ANNUAL PRODUCTION — 60 Billion CF

1.	LNG Liquefaction Plant Investment	\$16,000,000
2.	Capital Cost	
	Annual Principal	1,067,000
	6% Average Interest	480,000
	6% Tax Rate and Insurance	960,000
	Total	\$ 2,507,000
3.	Sale Gas	
	5.45¢/1000 CF X(60 X 10^9 CF)	\$ 3,270,000
4.	Cost of Liquid Nitrogen	\$10,608,000

$$\$0.0025/\text{lb} \times \frac{1.7 \text{ lb } LN_2}{\text{lb LNG}} \times \frac{41.6 \text{ lb}}{1000 \text{ CF}} \times (60 \times 10^9 \text{ CF})$$

5.	Annual Operating Expense	
	7% of \$16,000,000	\$ 1,120,000
6.	Power Credit	
	10,000 kW X 7200 hr X 0.005 kWhr	\$ 360,000
7.	Total Cost, After Power Credit	\$17,145,000
8.	Unit Cost of Vaporized LNG	28.6¢/1000 CF

2.1.2.5.3.2. Cold Gas Refrigeration

Present operation requires that rod seals be replaced every 2000 hours of operation. This requires shutdown for 4 man-hours per block. Other maintenance is light because of the relatively simple construction, minimal controls and piping, and moderate speed of operation. Full capacity is obtainable during the 300-day annual cycle of the other liquefaction system.

The operating cost structure is shown in Table 2.1.2.-9. Note that a 10-year straight-line amortization schedule is followed rather than the 15-year one of the other systems. It is felt that the reciprocating parts and bearings do not have the life capability of the rotary equipment used in other methods.

Table 2.1.2.-9. CALCULATION OF UNIT COST OF LNG
ANNUAL PRODUCTION – 750 Million CF

1.	LNG Liquefaction Plant Investment	$2,000,000
2.	Capital Cost (10-yr Amortization)	
	Annual Principal	200,000
	6% Average Interest	60,000
	6% Tax Rate and Insurance	120,000
	Total	$ 380,000
3.	Sales Gas	
	5.45¢/1000 CF X 750 X 10^6 CF	40,880
4.	Power Cost at $0.005/kWhr	64,505
5.	Annual Operating Expense	
	8% of $2,000,000	160,000
6.	Total Cost	645,380
7.	Unit Cost of LNG	86¢/1000 CF

The cost of LNG is 86.0¢ /1000 CF, which is the highest cost of any of the methods analyzed. It would seem that this system would be applicable in special situations such as recovery of boil-off gases that might otherwise be flared, as in a tanker with fuel-oil-burning engines and no provision for natural gas enrichment. However, even here, unless costs could be markedly reduced, it would be more economical to flare the gas than to recover it.

2.1.3. Storage Systems

Large volumes of natural gas can be stored in a liquefied state. As a liquid it will occupy approximately 1/600th of its gaseous volume under standard conditions. Since large-volume storage is desired for most applications, the internal tank pressure is advantageously limited to near-atmospheric for structural reasons. To maintain the natural gas as a liquid at atmospheric pressure, it is necessary to maintain the temperature at approximately −260°F; thus, the concept of refrigerated storage evolved.

Storage containers of interest for power generation will be on the order of 1,200,000 bbl (42 gal/bbl) near seaways where LNG is imported. Smaller containers for 100-MW turbine generators would be approximately one-sixth this size, assuming that they would store sufficient LNG for 30 days of rated capacity operation. The following descriptions will apply to tanks in this size range.

Two areas of interest have indicated a definite need for large-volume storage of LNG. In transporting gas from energy-rich to energy-poor areas in insulated tankers, the need for large storage depots at both the loading and the market ports is evident. In such base-load operations the storage facility is essentially a surge tank smoothing out the nonuniform base-load demand and fuel delivery operations. In this case, it is not necessary to preserve the LNG in the tank for long periods, because, as it is passed on to distribution within a short period of time, the higher heat influx resulting in greater boil-off rates can be tolerated.

In the second area, the continued growth and expansion of the natural gas industry has created the necessity for large-volume storage near metropolitan areas to meet the winter peak loads. In this case, LNG is stored for relatively long periods of time and used (i.e., vaporized and distributed) during only a few days of the winter. Consequently, heat influx must be held to a minimum

Whether the LNG plant is to be used for base-load operation or peak-shaving, a large portion of the cost is the storage facility. Also, the large volume of storage associated with LNG plants represents a high concentration of energy. It is for these reasons, i.e., economics and reliability, that recent research and development efforts have been concerned with storage concepts.

Conventional aboveground double-walled metal tanks have been used successfully for cryogenic service for several years. These tanks have generally been flat-bottomed, cylindrical, dome-roofed, double-walled tanks with an outer shell of carbon steel and an inner tank of 9% nickel steel or aluminum. Insulation between the two shells usually consists of perlite in a nitrogen or natural gas atmosphere.

Another form of cryogenic storage that has been used is the prestressed concrete tank. From the technology developed during the design and construction of hundreds of tanks and reservoirs for liquid storage, two aboveground double-walled prestressed concrete tanks were built for liquid oxygen service. These tanks have been in service for several years.

In attempts to reduce storage costs, several comprehensive research and development projects were carried out in the late 1950's and early 1960's. Among these were two projects investigating the feasibility of buried, prestressed concrete tanks. Both projects affirmed the technical feasibility of large-volume storage in such tanks. They also resulted in improved and more economical concepts for above- or belowground storage in prestressed concrete tanks.

Another concurrent project investigated the possibility of storing the LNG in frozen holes. In this method a pit is excavated in water-saturated soil by first freezing a ring of soil with refrigerant circulated through a series of buried, concentric pipes. Two major contemporary LNG facilities are using this method of storage.

Still another storage method receiving development attention is the mined-cavern concept. Such caverns, mined through vertical shafts in limestone, granite, chalk, shale, and dolomite, have been used for several years for liquefied petroleum gases. Theoretical, laboratory, and pilot plant studies have indicated that LNG storage in mined caverns is feasible.

LNG storage entails problems of compatibility and durability of the construction materials. The storage container must perform the dual functions of safely containing the liquid and minimizing boil-off. In some types, notably the cryogenic in-ground system (CIG), no attempt is made at insulating. The types of storage which have been developed and are under development are as follows:

1. Aboveground Metal Tanks
2. Cryogenic In-Ground Storage
3. Prestressed Concrete Tanks
4. Mined Cavern Storage

At the present time installations utilizing the first three concepts or planning application of them are in service, and are discussed in the following sections.

2. 1. 3. 1. Aboveground Double-Walled Storage

2. 1. 3. 1. 1. Metal

2. 1. 3. 1. 1. 1. Description and Operating Characteristics

Metal aboveground double-walled storage tanks are one of the most popular methods of storing LNG. These tanks are constructed with inner liners of 9% nickel steel or aluminum. At the present time, 41 aboveground double-walled LNG storage tanks are in service or under construction (Table 2.1.3. -1.).

Storage of LNG in aboveground metal tanks is accomplished in a vessel more fully described as a flat-bottomed double-walled dome-roofed tank. Figure 2.1.3-1 illustrates a typical double-walled flat-bottomed tank and indicates its principal components.

Experience with large cryogenic metallic tanks is not limited to the LNG tanks. Due to the space program and the rapidly expanding cryogenic field in general, over 350 similar types of tanks have been constructed for storage of liquid oxygen, nitrogen, argon, helium, and ethylene. Storage of these other cryogenic fluids has been accomplished not only in the double-walled flat-bottomed type of tank, but also in double-walled spheres, spheres suspended in cylindrical vessels, cylindrical tanks suspended within cylindrical tanks, and various other configurations of metal tanks. Selection of tank configuration is on the basis of requirements for storage capacity, operating pressure, external loading, and other factors dictating service conditions.

Table 2.1.3.-1. ABOVEGROUND DOUBLE-WALLED METAL LNG TANKS

No. of Tanks	Dimensions Diam x Ht (Inner and Outer)	Material (Inner Liner)	Capacity, Each		Owner	Location	Completion Date
			Liquid, 1000 bbl	Equiv Gas, 10^6 SCF			
3	142' D x 80'-3" H 148'-4" D x 84'-9" H	Aluminum	225	775	Phillips Petroleum	Kenai, Alaska	1969
2	129' D x 94'-6" H 135' D x 104'-6" H	9% Nickel	220	760	Tokyo Gas Co.	Tokyo, Japan	1969
2	142' D x 94'-6" H 152'-10" D x 104' H	9% Nickel	290	1000	Tokyo Gas Co.	Tokyo, Japan	1969
1	142' D x 105'-3" H 149' D x 109'-5" H	9% Nickel	300	1000	Boston Gas Co.	Boston, Mass.	1968
1	118' D x 94' H 128' D x 95'-2" H	9% Nickel	175	600	Northern & Central Gas Co. Ltd.	Hagar, Ontario	1968
2	138' D x 118' H 144' D x 123'-0" H	9% Nickel	300	1035	Esso-Libya	Libya	1968
1	118' D x 95'-11" H 128' D x 98'-1" H	Aluminum	175	600	Northwest Natural Gas Co.	Portland, Oreg.	1969
1	129' D x 130'-10" H 137' D x 135' H	9% Nickel	304	1050	Memphis Gas Division	Memphis, Tenn.	1966
1	118' D x 91' H 128' D x 94'-2" H	9% Nickel		600	Brooklyn Union Gas Co.	Greenpoint, N.Y.	1968
1	71'-6" D x 102'-3" H 81'-6" D x 105'-6" H	Aluminum	72	250	Wis. Natural Gas Co.	Oak Creek, Wis.	1965
1	111' D x 98' H 117' D x 106' H	9% Nickel	175	600	San Diego Gas & Electric Co.	Chula Vista, Calif.	1965
1	118' D x 91' H 128' D x 94'-2" H	9% Nickel	175	600	Alabama Gas Corp.	Birmingham, Ala.	1965
3	83'-6" D x 79'-6" H 89'-5" D x 87'-6" H	9% Nickel	76.5	264	Gaz de France	Le Havre, France	1964
5	95' D x 47' H 101' D x 52' H	Aluminum	60	200	North Thames Gas Board	Canvey Island, England	1963
1	80' D x 80'-2" H 86' D x 87'-5" H	9% Nickel	70	240	CAMEL	Arzew, Algeria	1963
2	80'D x 80'-2" H 86' D x 87'-5" H	Aluminum	70	240	CAMEL	Arzew, Algeria	1963
2	50' D x 46' H 56' D x 52' H	Aluminum	16	55	North Thames Gas Board	Canvey Island, England	1958
1	67' D x 56' H 73' D x 61'-6" H	Aluminum	35	120	Constock Liquid Methane Corp.	Lake Charles, La.	1958

Table 2.1.3-1, Cont. ABOVEGROUND DOUBLE-WALLED METAL LNG TANKS

No. of Tanks	Dimensions, Diam x Ht (Inner and Outer)	Material (Inner Liner)	Capacity, 1000 bbl	Owner	Location	Completion Date
1	142' D x 105' H 149' D x 109'-5" H	Aluminum	290	Transcontinental Gas Pipe Line Co.	Carlstadt, N.J.	1969
1	78' D x 93'-5" H 88' D x 96'-10" H	Aluminum	78	Northern States Power Co.	Eau Claire, Wis.	1969
1	63' D x 70'-9" H 73' D x 74' H	Aluminum	38	Northern States Power Co.	La Crosse, Wis.	1969
1	142' D x 112' H 151' D x 114'-9" H	9% Nickel	300	Quebec Natural Gas Co.	Montreal, Canada	1969
1	186' D x 60'-6" H 194'-10" D x 65' H	9% Nickel	290	Lowell Gas Co.	Lowell, Mass.	1969
1	154' D x 111'-6" H 164' D x 116'-3" H	Aluminum	350	San Diego Gas & Electric Co.	Chula Vista, Calif.	1970
1	94' D x 41' H 101' D x 52' H	Aluminum	60	North Thames Gas Board	Canvey Island, England	1967
1	95' D x 62' H 101' D x 65' H	Aluminum	75	East Midlands Gas Board	Ambergate, England	1969
2	161' D x 88' H 171' D x 93'-6" H	9% Nickel	300	SNAM	La Spezia, Italy	1969

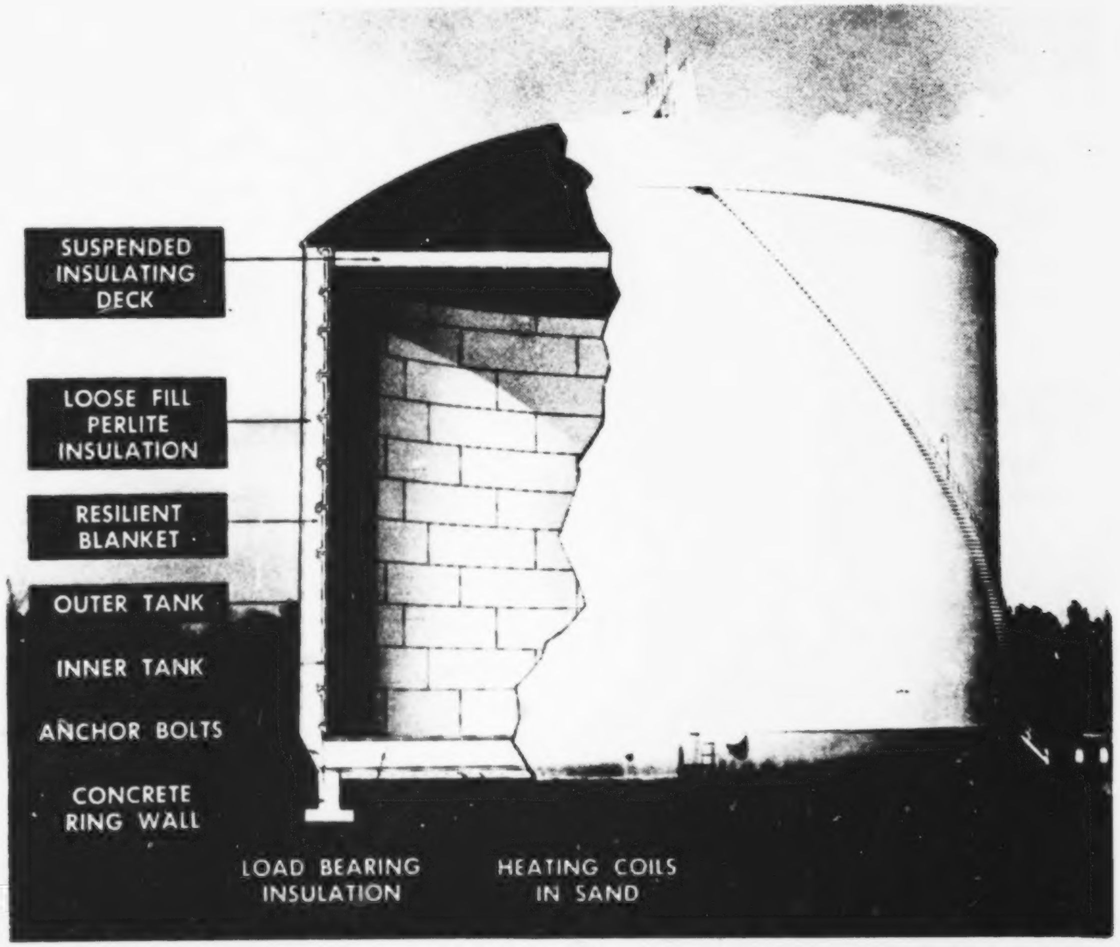

SUSPENDED INSULATING DECK
LOOSE FILL PERLITE INSULATION
RESILIENT BLANKET
OUTER TANK
INNER TANK
ANCHOR BOLTS
CONCRETE RING WALL
LOAD BEARING INSULATION
HEATING COILS IN SAND

Figure 2.1.3.-1. TYPICAL ABOVEGROUND DOUBLE-WALLED TANK WITH ARTIST'S CUTAWAY SECTION [30]

The double-walled flat-bottomed tank is really a tank within a tank, with the annular space between the two filled with insulating material (Figure 2.1.3.-L). The inner tank, in contact with the LNG, must be made of materials of proper strength for the design loadings at the temperature of the stored liquid.

The outer tank serves the primary purpose of containing the insulating material that surrounds the inner tank. The outer tank provides the best obtainable means of protection for the insulation system against external forces, such as fire, impact, and particularly, weather and moisture. The annular space is vaportight; in it a dry atmosphere is maintained. Connections to the inner tank must pass through the outer tank, and are of a special design to permit differential movements due to temperature and, at the same time, maintain the leaktight integrity of the overall system.

The approach to the design of a double-walled metal LNG tank is on a more conservative basis than is used for the design of now commonplace tanks storing oil, water, and other liquids. The approach is similar, and many of the same rules and standards are used, but it takes into account what LNG requires due to its low temperature and other storage conditions. If the designer and constructor follow the now-established LNG tank design criteria, the resulting tank should be safer and more adequate than any of the many thousands of tanks already in oil and water service, where similar design approaches have been taken, but on the basis of the particular liquid.

The metal for constructing the inner tank is one that retains its strength and ductility throughout the range of temperatures possible in the tank. The material must be particularly notch-tough for the important and more highly stressed components. Aluminum, stainless steel, and 9% nickel steel are known and have proved their ability to remain notch-tough beyond the extremely low temperatures of LNG. Where a material exhibits a transition from ductile to brittle behavior as the temperature decreases, acceptance is based upon rigid tests that the material would normally be expected to pass.

Site requirements for the location of aboveground double-walled tanks make it possible to select any site of sufficient area. Soil conditions may require piling. Concrete ringwalls are most commonly used on soils with

2000-3000 lb/sq ft bearing. The tanks can be decreased in height and increased in diameter to accommodate poor soils; however, optimum proportions for the most economical cost of storage are achieved by utilizing high tanks on good bearing soils of 4000 lb/sq ft.

The tanks are provided with dikes having a volumetric capacity of up to 150% of the tank storage capacity, except where protection is provided by the topography, such as a natural hollow which could serve as a ponding area in the event of a liquid spill. The dikes are usually constructed of earth and designed to be liquidtight and to withstand the full hydraulic head and the force of any major spillage.

If the tank foundation is of the ringwall type, then the outer bottom rests on a gravel base directly upon the soil. Because there is a certain amount of heat migration through the bottom of the tank, heating coils are installed under the bottom to prevent any soil freezing and consequent frost heaving. In the case of a foundation supported on piles, a minimum 18-in. air space is provided between the ground surface and the bottom of the pile cap to allow for the circulation of air, which eliminates the need for any additional heat source, such as coils.

In LNG and cryogenic storage systems, insulation is a primary factor. Proper insulation is the key to successful and economical storage.

The outer tank of the double-walled tank serves as a vapor and protective barrier for the insulation system. With LNG tanks, where the storage temperature is below the atmospheric dew point, moisture would migrate into an unprotected insulation, where it would condense. Since temperatures within the insulation are below 30°F, the moisture would freeze and ice would damage the insulation, causing a significant loss of insulating efficiency and, eventually, destroying the insulation. For LNG tanks this is particularly serious because they stand cold for long periods without warming up, affording no opportunity for reversal of moisture migration. The outer steel shell of the double-walled LNG tank represents the best possible vapor barrier obtainable against this migration of moisture. It permits a gastight insulation space where a dry atmosphere can be maintained.

It is not economical to design the outer tank to withstand the maximum external pressure that could arise from variations in atmospheric conditions; therefore, a slight positive pressure is maintained in the space with a dry gas that would not liquefy at the storage temperature. Nitrogen or methane is normally used. The method now being used is to suspend an insulated roof from the outer tank roof and permit boil-off gas to provide the necessary positive pressure. Such a system is illustrated in Figure 2.1.3.-1.

The insulation material commonly used in double-walled LNG tanks is an inert, inorganic, noncombustible, granular material commonly called "perlite." Perlite is a type of naturally occurring volcanic glass rock containing trapped water. When heated in a kiln above 1500°F, the crude perlite particles expand and turn white. The trapped water vaporizes to form microscopic cells or voids in the heat-softened glass. The resulting sealed-cell honeycomb structure accounts for a very lightweight and excellent thermal insulation material. The crude perlite ores are carefully selected and graded according to size prior to expansion so as to produce desired gradations and densities when expanded. The expanded perlite normally used has densities of 3 to 5 lb/cu ft.

A reasonable value used for the "apparent mean thermal conductivity" or "K" value, in Btu per in. of thickness per sq ft per hr per °F of the perlite system is 0.24 for a mean storage temperature of −260°F with an 80°F ambient temperature. The installed thickness of the perlite governs the overall rate of heat influx through the system. The usual standard is based upon at least 3 feet of thickness. The amount of heat influx into the tank determines the more commonly used factor of "boil-off," which refers to the percentage of the tank's contents which will boil off in a given period of time under specified ambient temperature conditions.

For example, a 1 billion CF capacity LNG tank (290,000-bbl liquid capacity) with 3 feet of perlite insulation would have a boil-off rate of approximately 0.05%/day when the tank is full of liquid, on a day when the average ambient temperature is 80°F.

Figures 2.1.3.-2 and 2.1.3.-3 illustrate heat leak and cooldown for double-walled LNG tanks with the inner tank made from 9% nickel, sidewalls and roof insulated with 3 feet of loose fill perlite, and the base insulated with 3 feet of Foamglas or equivalent load-bearing insulation such as cored perlite concrete block with loose fill perlite.

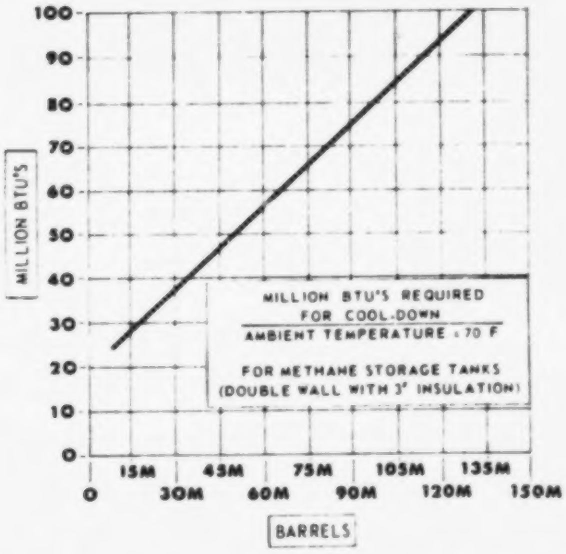


Figure 21.3.-2. COOLDOWN OF DOUBLE-WALLED METHANE STORAGE TANKS WITH 3 ft INSULATION[5]

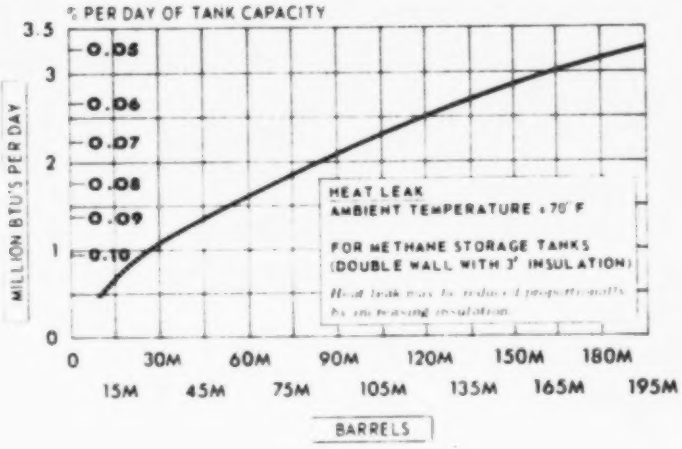


Figure 2.1 3.-3. HEAT LEAK OF DOUBLE-WALLED METHANE STORAGE TANKS WITH 3 ft INSULATION[5]

Perhaps a more meaningful expression of a system's thermal efficiency would be in terms of average boil-off, based upon a year's cycle of operations, starting from an empty tank to a full tank and taking into account daily ambient temperature variations throughout the year. This rate would be substantially lower than the maximum daily rate indicated above, but it is a complex figure to determine.

Heat influx and boil-off rates of double-walled tanks can be improved or furnished to meet almost any desired rate. Additional improvement can be achieved by increasing the thickness, which on some existing installations has been up to 5 feet. A vacuum could be introduced into the insulation space; this would reduce gas conduction and convection transfer to negligible values and, thus, reduce the thermal conductivity. For large LNG tanks, however, the use of vacuum insulation systems would not be economically justified.

Other types of insulation material could be utilized in the annular space, such as Fiberglas, polyurethane foams, Foamglas, or even mineral wools. Some of these materials have better K values in comparison to perlite alone; however, when taking into consideration the cost of these insulating materials and their installation, the perlite system has a substantial cost advantage over all others. The installation of perlite is a simple and economical operation, as the ore is expanded at the tank site and blown directly into the insulation space without any manual handling.

Perlite, by itself, is not a resilient material and compacts in use. Because of tank operation cycles and changes in storage and ambient temperatures, differential movement occurs between the inner and outer tanks. This movement would further tend to compact the loose perlite material. Therefore, means must be provided in the insulation system to control compaction and to accommodate differential movement between the inner and outer shells. This aspect of the perlite double-walled insulation system is covered by various U.S. and foreign patents held by the Chicago Bridge & Iron Company. The system most commonly used utilizes the "Fiberglas resilient blanket," which is a blanket completely covering the external surfaces of the inner tank. The blanket varies in thickness, depending on the differential movement expected, and is sufficiently resilient to maintain a firm bearing against the loose perlite while taking up the differential movement between the inner tank and the wall of perlite. Such provisions are an important factor, and perhaps the key to an economical and successful double-walled tank insulation system.

The problem of insulating the space between the tank bottom is different than for the annular spaces between the walls and roof. In this area, a load-bearing type of insulation is used. Earlier methane tanks used the block material known as Foamglas, which proved satisfactory except that it was costly and somewhat fragile. It was easily damaged during installation and during subsequent periods when the tank erection work was under way. There were problems in keeping the material dry during construction.

A recent development, now being used on LNG tank construction, is the Chicago Bridge & Iron Company's system of cored perlite concrete blocks. These are large precast blocks of lightweight perlite aggregate concrete, which alone has desirable insulating properties. The blocks are honeycombed with open vertical cores representing between 65 and 75 % of their surface area. These open cores are later filled with the same type of loose perlite used in the tank's annular space. The blocks have sufficient strength to support the weight of the tank and its contents. This new system has improved the thermal properties of the bottom insulation of double-walled LNG tanks, and has also made possible a significant improvement in costs.

Other basic design considerations of the double-walled system are a) to achieve an insulation system with the same life as the metal tank, b) to furnish an insulation system that could not be damaged by unusual external forces such as fire, weather, or impact, and finally c) to make possible the safest, most economical type of metal tank yet constructed for cryogenic liquid storage.

The double-walled tank does not impose any unique operating conditions upon the overall LNG plant operating picture. Operation of the tanks is considered foolproof, as they incorporate a complete range of adequate and proved safety and operating accessories. Over- or underpressuring of the tank is prevented by duplicate systems of pressure and vacuum vents to ensure satisfactory operation in the event of malfunction of any one. This duplication also permits routine maintenance or inspection of one unit while the other unit remains in service to provide the intended protection.

Overfilling is prevented by a tank overflow that would lead the liquid back into the plant or flare system. Prior to reaching the overflow point, a system of alarms and warnings would call the operator's attention and automatically cut off the flow into the tank.

After initial cooldown, atmospheric pressure and temperature changes do not influence tank operation significantly, so subsequent day-to-day operation of an LNG storage tank is really no different than what is considered normal operation for more conventional oil and water storage tanks.

Maintenance costs of double-walled LNG tanks are considered very low, and no different than they would be for conventional tanks storing other liquids. The only significant maintenance cost of steel tanks is periodic external painting.

2.1.3.1.1.2. Investment Costs

For the case of aboveground double-walled metal storage tanks, we can draw on the experience gained from some 37 units, which will all be built by the end of 1969. These tanks of 16,000-bbl capacity or larger have been fabricated of either aluminum or 9% nickel steel inner tanks in some seven countries.

In presenting the following investment costs certain trends in costs and significant cost parameters should be noted. In the past several years there has been a general decrease in the cost of metal tankage whereas other types of tankage have increased in cost. This is in part due to the advances in construction techniques gained from the experience of constructing the 37-tank units.

Frequently it is assumed that any type of storage is equally feasible and competitive at any storage site. From some of the basic site requirements that are given in Table 2.1.3.-2., it can be seen that certain sites could impose physical as well as economic restrictions on any one storage system that would preclude its use. In terms of the aboveground metal tank, consideration should be given to special foundation costs and land requirements.

Special foundation costs for an aboveground tank could add considerably to the total investment cost. If the bearing capacity of the soil is below the 2000-4000 lb/sq ft value, piles will be required. If the area is extremely marshy as it might be near the sea coast, the soil may not be able to bear the containment levee. Required support for the levee would also have to be added to the investment costs. The minimum tank spacing for aboveground tanks is greater than those classed as belowground. Therefore, the additional incremental land costs must be evaluated and debited against the aboveground storage tanks. Thus, in evaluating a given site for the lowest investment cost, many factors such as soil type and bearing strength, Table 2. 1. 3. -2. BASIC SITE REQUIREMENTS FOR VARIOUS TYPES OF LNG STORAGE [45]

	Frozen Holes, In- Ground Containers	In-Ground Tanks	Aboveground Tanks
Distance from Edge of Tank to Property Line	100 ft	100 ft	200 ft
Distance Between Containers	10 ft*	10 ft	1/4 Sum of diameter of adjacent tanks
Containment Dikes Required	No	No	Yes
Desired Location of Permanent Water Table	Near surface	Below tank bottom	Not critical
Required Soil Bearing Without Pile Foundation	None	2000-4000 lb/sq ft	2000-4000 lb/sq ft
Impervious Bottom Strata	Required	Not reqd	Not reqd
Maximum Size in One Container	Upper limit not yet defined		Approximately 400, 000 bbl

* This is the recommended safety spacing. However, for frozen holes this would probably be insufficient for structural strength.

groundwater table, physical properties of subsurface, strata, and land costs will be of importance.

The investment cost would, therefore, appear to be unique for each site. Even so, it is possible to create generalized cost data for comparative purposes.

One such estimate is shown in Figure 2. 1. 3. -4 for storage capacities from 0. 25 to 10 billion SCF of LNG equivalent.[65] The estimates are for both single- and multiple-tank groups. It is assumed that the maximum current single-tank capacity is 400, 000 bbl.

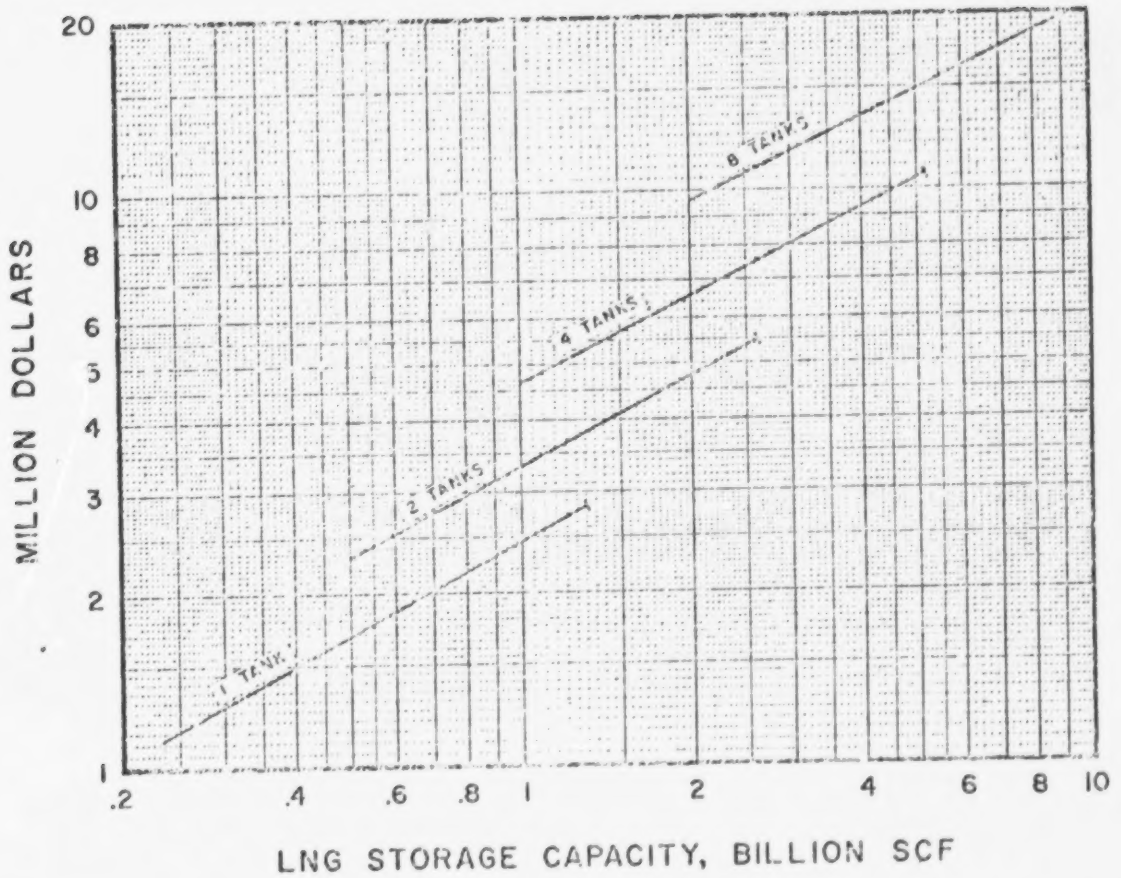


Figure 2.1.3.-4. LIQUEFIED NATURAL GAS STORAGE CONSTRUCTION COSTS* USING ABOVEGROUND METAL TANKS [65]

* Note: Actual costs may vary considerably. Data are, therefore, intended for use only as a guideline.

The inner tank in this estimate is constructed of 9% nickel steel. The outer carbon steel tank retains a 3 ft thickness of granular perlite insulation surrounding the inner tank. It is assumed that the heat leakage into the tank would be equal to 0.08%/day vaporization of LNG from the filled tank. This is an average figure, as the loss should be relatively greater for a small tank and less for a large tank.

The estimated costs include the facilities required to make the storage tank a complete working installation, such as a foundation, a diked enclosure having a capacity 1.5 times the volume of the storage tank (even though 1.0 times is all that is required by the NFPA Standard No. 59A for the Storage and Handling of Liquefied Natural Gas at Utility Gas Plants), instrumentation, safety devices and controls, piping and insulation, pipe carriers, and general service facilities on a prorated basis. The costs for all piping between the storage tanks and the liquefaction and sendout units are charged to the storage area. An allowance is also made for fire protection around the base of the storage tank.

It is assumed that the location for a proposed LNG installation is in a poor, rather than good, soil for load-bearing purposes, and that piled foundations are required. Costs are based on a typical condition for which 50-ft-long piles would be adequate. If good foundation conditions are found, piling can be omitted and the storage area cost indicated in Figure 2. 1. 3. -4 can be reduced by $200/million SCF of LNG storage capacity.

The costs used are those of the 3rd quarter of 1964. They may be adjusted for current costs by using the "Equipment Cost Index," compiled quarterly by Marshall and Stevens of Los Angeles, California. This index is quoted in each issue of Chemical Engineering and is also in the Engineering News-Record, both published by McGraw-Hill. The value of this index was approximately 242 in the 3rd quarter of 1964 (1926 = 100), and for the 3rd quarter of 1968 was 274. 8.

The costs given above apply to the facilities necessary to make a storage tank a complete working installation. They, therefore, represent construction costs as contrasted with investment costs. The addition of applicable charges for land, and the cost of money would bring these construction costs in line with other published investment costs.

The most recently available estimated investment cost data for a specific site and storage volume are the tanks to be constructed for Tennessee Gas Transmission Co. at Hopkinton, Massachusetts. The three tanks to be constructed will have a total capacity of approximately 870, 000 bbl: each being 290, 000 bbl. The estimated investment cost of the tanks of standard design is given in Table 2. 1. 3. -3. Comparing this cost estimate to the more generalized estimate in Figure 2. 1. 3. -4 shows that the Hopkinton cost, minus land and interest of $9, 262, 800, is 16. 6% greater than the updated and interpolated cost of $7, 722, 000. The difference is within the expected variation between a specific and generalized cost estimate.

Table 2. 1. 3. -3. LNG TANK COST FOR TENNESSEE GAS TRANSMISSION CO.

Item	Unit Cost, $	Amount, $	Total, $
Land	2000	102, 000	
Site Investigation		11, 000	
Improvements		63, 000	
Storage Tanks			
Double-walled cryogenic above-ground tanks C/W* foundations, heating coil, insulation, painting, etc.		5, 900, 000	
Dikes C/W* Roadways, Walkways, etc.		440, 000	
Sump Pumps for Dike Area — Installed		24, 000	
Electrical		64, 000	
Instruments and Controls		86, 000	
Building		25, 500	
Blowers — Vapor Return		82, 500	
Pipes, Valves, Fittings		325, 000	
Fire Protection		62, 500	
Installation		618, 500	
Total			7, 804, 000
Contingencies at 10%			780, 400
Overhead at 10%			780, 400
Interest at 9%			702, 360
Total			10, 067, 160

* C/W = Cost with.

2.1.3.1.1.3. Operating Costs

The operating costs for an aboveground double-walled metal storage tank include charges for operating the belowground electric heaters (where applicable), maintenance, operating labor, operating utilities, operating supplies, and liquefaction cost of the heat leak gas.

The operating costs based on 1 billion SCF of LNG equivalent are given in Table 2.1.3.-4. These costs are devised from the costs of a typical cascade-type liquefaction plant designed for medium natural gas. The assumed feed gas pressure is 300 psig which is raised to 640 psig in the feed gas booster compressor. The heat leakage rate is 0.08%/day on a full tank basis. The heat leak gas is considered to be fuel or sendout gas. For liquefaction the fuel gas charge is \$0.26/1000 CF, and the electricity charge is \$0.01/kWhr. The plant operating staff consists of two operators per shift with one-third of their time chargeable to the storage unit.

Table 2.1.3.-4. ANNUAL OPERATING COSTS FOR A 1 BILLION SCF EQUIVALENT STORAGE FACILITY FOR METAL TANKS

1.	Belowground Heaters, 300,000 kWhr/yr at \$0.01/kWhr		\$ 3,000
2.	Maintenance		\$13,965
3.	Operating Labor		\$26,667
4.	Operating Utilities, Electric		
	Blowers Lights Pumps	1 million kWhr/yr at \$0.01/kWhr	\$10,000
5.	Operating Supplies		
	Inert Gas Lubricating Oil Gland Cooling Water		\$ 5,000
6.	Liquefaction Cost of Heat Leak Gas, 292 X 10^6 CF at \$70/10^6 CF		\$20,440
		Total	\$79,072

2. 1. 3. 1. 2. Concrete Tank Storage Systems

2. 1. 3. 1. 2. 1. Description and Operating Characteristics

Prestressed concrete tanks may be built above- or belowgrade. When built abovegrade they may be of either single-walled or double-walled construction. If single wall, the insulation may be internal or external. When perlite or a similar material is used for external insulation, an outer wall is usually constructed to keep the insulation dry. This outer wall, of double-walled construction, may be made of carbon steel or prestressed concrete. If the outer shell is prestressed concrete, it also serves as a safety wall, but at some cost disadvantage.

The operating characteristics are similar to those of an aboveground metal tank.

2. 1. 3. 1. 2. 2. Investment Costs

The only difference in structural components between aboveground concrete and metal tanks is the inner vessel. It is noted that the diameter of the concrete shell is greater than a corresponding metal shell because of the comparative thickness of the concrete. For equal insulation thickness, slightly more insulation and a larger outer shell is required with concrete. This cost increment is also reflected in foundations, etc.

On the same basis of estimating construction costs as in Section 2. 1. 3. 1. 1. 2., the additional cost due to substituting a prestressed concrete inner shell for a metal inner shell updated to the 3rd quarter of 1968 would be –

Storage Capacity, 1000 bbl	Additional Cost, $
200	170, 000
300	256, 000
450	454, 000

The costs given are for single-tank construction.

2.1.3.1.2.3. Operating Costs

Operating costs for a concrete storage tank would not be significantly greater than for a comparable metal tank (Section 2.1.3.1.1.3.).

2.1.3.1.3. Vacuum-Insulated Storage

2.1.3.1.3.1. Description and Operating Characteristics

In general, vacuum insulation is used in relatively small cryogenic containers because it is the most efficient and compact type of insulation. However, it has been used on vessels up to 105,000 gal where weight and space saving have been critical factors. The difficulty of ensuring a vacuum and of maintaining it has prevented its wider use. A barge-mounted 105,000-gal tank is being used in the space program to transport liquid-oxygen (LOX). The boil-off of this tank is 0.15%/day.

LOX railcars with a capacity of 31,900 gal gross and 30,000 gal net are used. These are vacuum-insulated. In most of these designs, perlite powder fills in the vacuum space. The latest LNG railcars have a capacity of 33,000 gal net. This car is 76 feet long and 10 feet in diameter. The perlite-filled annular space varies between 5 and 8 in. It is interesting to contrast this thickness with the larger nonevacuated tanks that have perlite walls up to 5 feet thick.

The basic principle of the vacuum is to minimize the conduction paths between the cold inner vessel and the warm outer vessel. By removing the air molecules, heat can be transmitted only by radiation and by conduction between the light support structures.

There are three general types of vacuum insulation, typified by the radiation barrier employed:

1. No shields or other radiation barriers
2. Powder or fine-grained particles radiation shields
3. Multilayer laminar shields

In the first case, a vacuum of at least 10 microns is used (a micron is equivalent to the pressure of a column of mercury one-millionth of a meter high). Since an atmosphere has a pressure of 0.76 meters of mercury, the vacuum required for good insulation is approximately one-millionth of atmosphere pressure. Above this pressure, conductivity increases rapidly.

The use of perlite or other powders in the evacuated space acts to reduce the ability of radiant energy to flow from the relatively warm outside wall to the cold inner wall. This is a surface effect. Perlite is an ash that is heated until the entrapped water forms steam, expands, and creates a thin-walled shell that cracks, permitting the steam to escape. The expanded perlite is then screened to obtain the desired range of product diameter. Figure 2.1.3.-5 shows the variation of thermal conductivity of various powders under various vacuums.

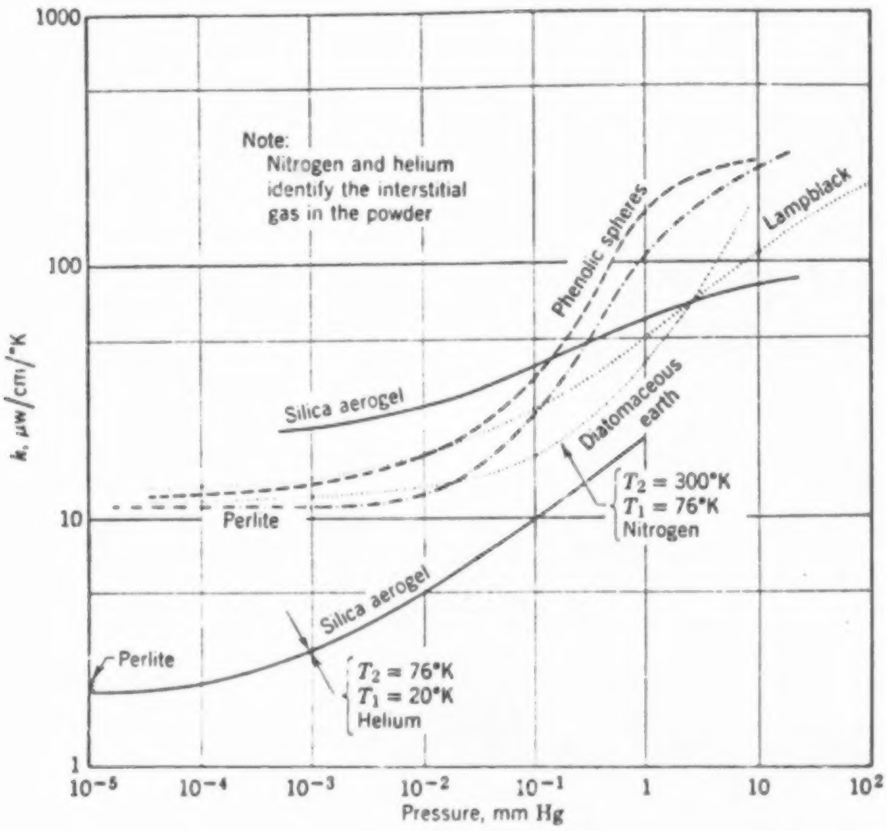


Figure 2.1.3.-5. APPARENT MEAN THERMAL CONDUCTIVITIES OF SEVERAL POWDERS [71]

Another feature of the powder insulation is that it reduces the need for structural support to separate the outer and inner tanks. The contact area between the two tanks is large, but the extreme length of the path, due to the small diameter of the particles and the low conductivity of the perlite, results in a low overall thermal conductivity.

The third type of vacuum insulation, multilayer, is often called "super insulation." It consists of alternate layers, as many as 20 per cm, of annular space, of highly refective foils, and insulating fibers. The foils may be aluminum or copper, and the insulators are glass fiber or nylon mesh. Aluminized mylar is also used. Figure 2.13.-6 illustrates the performance of this system.

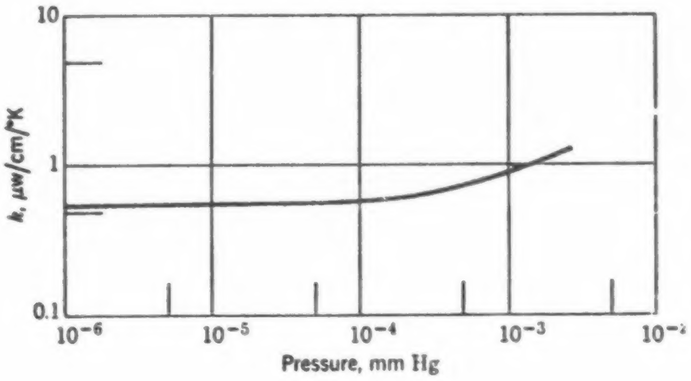


Figure 2.13.-6. DEPENDENCE OF APPARENT MEAN THERMAL CONDUCTIVITY OF MULTIPLE-LAYER INSULATION ON INTERSTITIAL NITROGEN GAS PRESSURE (Boundary Temperatures, 300° and 76°K)[71]

It should be noted that at 1 micron (10^{-3} mm Hg), this system is approximately 10 times better than the perlite system over the temperature range of interest: 300° and 76°K (LNG is at 111°K).

Vacuum containers may be of two types: permanent or powered. The permanent vacuum units are sealed after reaching the desired vacuum and must maintain their vacuum throughout their life. The "powered" system uses a vacuum pump to periodically lower the pressure to the desired level. The latter units do not require as careful a welding and cleaning procedure and are more suitable for large tanks.

Vacuum insulation would not be considered for use in base-load operations. This is because there is no demand for the extremely low boil-off rates attainable with this type of insulation.

2.1.3.1.3.2. Investment Costs

No investment costs have been prepared for this storage system as it is not applicable to base-load operations.

2.1.3.1.3.3. Operating Costs

No operating costs have been prepared for this storage system as it is not applicable to base-load operations.

2.1.3.2. In-Ground Storage

2.1.3.2.1. Concrete

2.1.3.2.1.1. Description and Operating Characteristics

Prestressed concrete tanks have been tested for LNG storage, but none have been built belowground. A single tank is being built aboveground at Barcelona, Spain, to receive tanker-delivered Libyan LNG. However, during the past 25 years, hundreds of prestressed concrete tanks and reservoirs have been designed and constructed. Their most extensive use has been for liquid storage tanks and reservoirs, both on the surface or belowground; techniques for storing liquids in such tanks seem to be well established. Included among these tanks have been two aboveground double-walled prestressed concrete tanks that have been in use for several years.

The Texas Eastern Transmission Corporation, working through Battelle Memorial Institute, developed a tank with special lining materials. A prototype of this development was built by Texas Eastern and tested by Battelle; it held liquid nitrogen for 15 months. This construction technique is embodied in the large-volume tank now under construction on Staten Island, N.Y. A similar method was investigated in a research project under the joint sponsorship of the American Gas Association and a group of companies. The tank was built by The Preload Company and tested by the Institute of Gas Technology.

A prestressed tank is designed and constructed so that the ring forces in a completely filled tank do not cause any circumferential tensile stresses in the concrete. To do this, high-tensile wires stressed on the order of 150,000 psi are wrapped around the concrete core. Figure 2.1.3.-7 shows the details of the tank construction. Vertical prestressing can be used to control the magnitude or to eliminate vertical bending stresses in the concrete wall.

Prestressed concrete tanks may be constructed aboveground, at grade, belowground, or partially belowground, depending on the site conditions or other factors. For tanks sited below grade, excavation and control of water follow standard procedures. Aboveground tanks may be either of the single-shell or dual-shell type, depending primarily on insulation requirements.

Floor construction can be adapted to meet special needs. The insulation may be placed over a prepared subgrade or subfloor with a 1/8 or 3/16-in. 9% nickel or stainless steel liner placed over the insulation. The insulation is then overlaid with a 2-in. lean concrete topping slab. An alternative method may also be used: Insulation may be placed over a prepared subgrade or subfloor and then the lean concrete topping applied. A carbon steel floor may then be installed, over which is placed a prestressed concrete floor. A third alternative for floor construction would be an integral insulation-liquid barrier arrangement placed over the tank floor and held down by concrete ballast. Regardless of which method is used, suitable means of connecting the floor system to the wall must be provided to permit contraction of the floor during initial cooling. This in turn will permit the walls to contract without undue restraint at the base.

With any of these methods of construction, an impermeable barrier, usually metal, is used on the exterior surface of the wall. The insulation and impermeable barrier may, however, be placed inside the concrete for protection from deterioration and mechanical damage. Walls may be vertically prestressed to control the stresses due to temperature gradients and the bending caused by the edge restraint of the cylindrical wall.

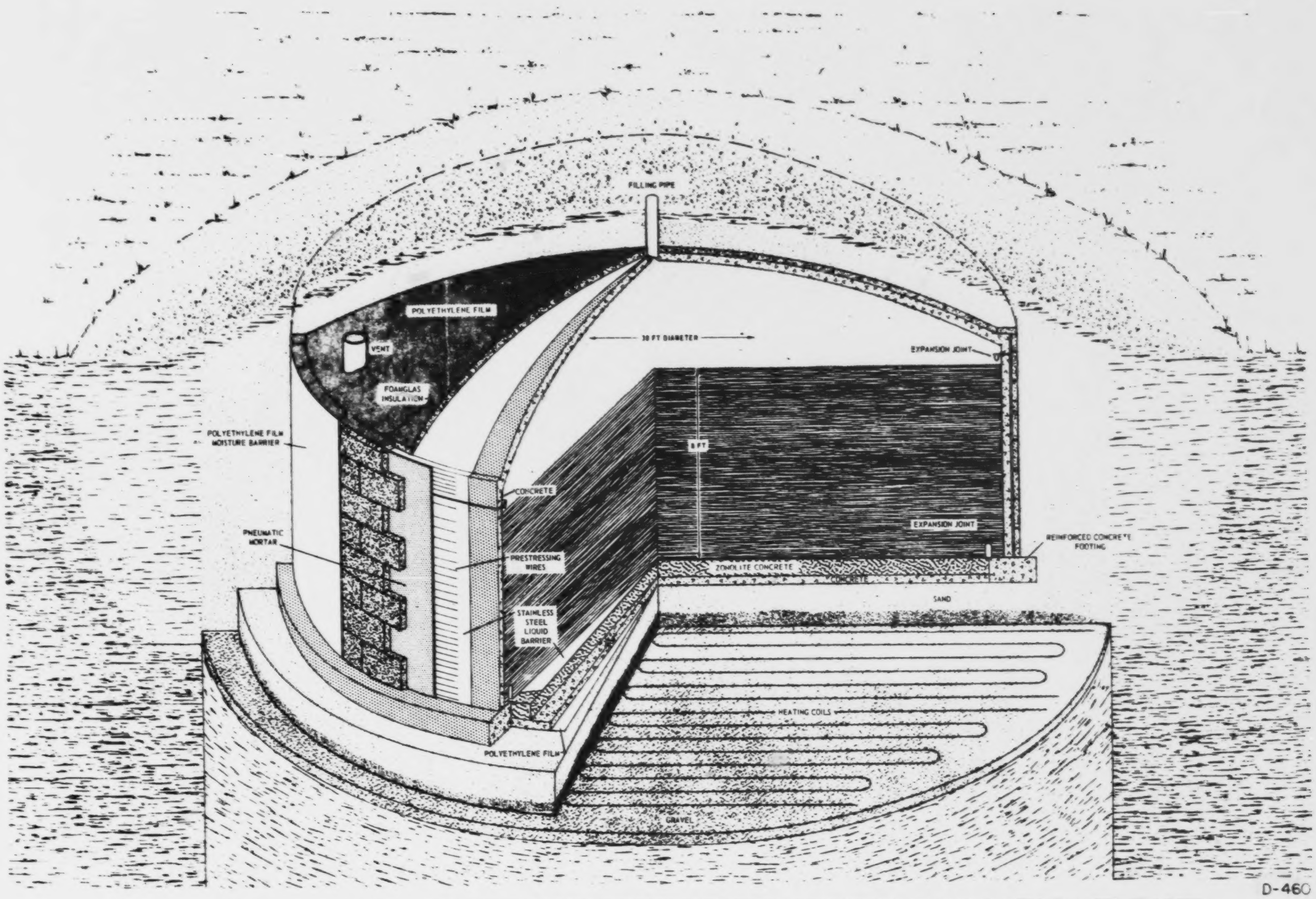

FILLING PIPE
POLYETHYLENE FILM
VENT
FOAMGLAS INSULATION
30 FT DIAMETER
EXPANSION JOINT
POLYETHYLENE FILM MOISTURE BARRIER
8 FT
CONCRETE
PNEUMATIC MORTAR
EXPANSION JOINT
REINFORCED CONCRETE FOOTING
PRESTRESSING WIRES
ZONOLITE CONCRETE
CONCRETE
SAND
STAINLESS STEEL LIQUID BARRIER
HEATING COILS
POLYETHYLENE FILM
GRAVEL
D-460

Figure 2.13.-7.BELOWGROUND PRESTRESSED CONCRETE LNG STORAGE TANK

Either metal or concrete roofs can be used. Although interior supports are acceptable, free-spanning domes are preferred. Roofs have been constructed in diameters up to 300 feet. The exterior roof surface is normally covered with either a metallic barrier or a polyester or epoxy-glass-reinforced coating.

One of the most important factors in construction of belowground tanks is the type and character of the soil, and the subsurface and foundation conditions. It usually is preferable to construct buried tanks above the water table, if at all possible. The dewatering requirements and other structural problems imposed by going below the water table usually eliminate the economies of belowground construction. Further precautions must be taken in frost-heaving soils. The tank must be protected against pressures created by ice lens formations. This may be done with heating elements placed below and around the tank or with below-tank heating elements and a frozen protective ring of select material around the tank.

The heat influx for belowground tanks is influenced greatly by whether or not the tank is located in a frost-heaving soil. If it is, and heating coils are provided, then a constant heat source at the temperature of 32°F exists at the heating coil approximately 3-5 feet from the tank. With this condition it is possible to accurately estimate the heat influx into the tank and provide the required insulation to meet the desired boil-off rate.

If the site is in non-frost-heaving soils and the soil is allowed to freeze, as the 32°F isotherm moves away from the tank, the heat influx diminishes.

The primary advantage of prestressed concrete tanks is that they can be built belowground, if the site conditions permit, or above grade. A belowground tank has at least a 100°F lower temperature gradient than a comparable aboveground tank, even with heating cables or coils. If no heating cables or coils are required, the advantages are even greater. In a relatively dry sand, after 1 year of operation the frozen sand creates an insulation equivalent to about 5 in. of urethane

A buried tank cannot spill its contents in the event of complete rupture or failure, even into a dike area. Seepage through the frozen ground is inhibited. In the event of fire, the burning surface would be limited to the tank area. Also public relations are enhanced by a buried tank: Since it is out of sight, it is out of mind.

The use and cost of buried prestressed concrete tanks is affected greatly by the subsurface conditions. This frequently can impose economic burdens on this type of construction. Repair of insulation is awkward for belowground tanks with the insulation on the exterior of the wall. This is obviously not a disadvantage for tanks insulated internally.

2. 1. 3. 2. 1. 2. Investment Costs

For the case of belowground concrete storage vessels, we can only draw on the experience of two demonstration-model-sized tanks and the 600,000-bbl tank being built by Texas Eastern Transmission Corp. at Staten Island, New York. Unfortunately, current investment cost data are not available for this complete tank facility. Generalized cost estimates must therefore be presented.

Estimated costs[65] for prestressed concrete tanks have been provided by the Preload Company for capacities of 100,000 to 450,000 bbl. Additional amounts are included to cover the costs of auxiliaries needed for complete storage facilities.

These costs include allowances for site improvements, electrical supply, instruments, safety and control devices, piping, foundation heating cables, and fire protection with costs of general service facilities prorated. Foamed-block insulation, applied to the outside walls to limit heat flow into the tank to that equivalent to vaporization of 0.1% of the full tank content per day, has been included.

This then puts the costs on the same basis as those presented in Section 2. 1. 3. 1. 1. 2; that is, the costs are those necessary to make the storage tank a complete working installation. They do not, however, include the cost of land. It should also be noted that the costs are based upon the water table being below the base of the tank. If this were not the case, the cost of artificially lowering and maintaining the lowered water table would have to be included in investment and operating costs.

The estimated costs of complete facilities incorporating belowground concrete tanks, backfilled and covered with earth, updated to the 3rd quarter of 1968 are shown below.

Storage Capacity		Cost of Storage Facility,
1000 bbl	10^9 SCF of NG	$
100	0.33	1,363,000
200	0.67	2,158,000
300	1.02	2,840,000
450	1.54	3,522,000

2.1.3.2.1.3. Operating Costs

The operating cost for a belowground concrete tank include charges for operating the belowground electric heaters, maintenance, operating labor, operating utilities, operating supplies, and liquefaction cost of the heat leak gas (Table 2.1.3.-5.).

These costs are derived on the same basis as those for the metal tank Section 2.1.3.1.1.3. For purposes of this estimate the operating costs of the belowground heaters, maintenance, operating labor are considered equal to those of the metal tank. Operating utilities, liquefaction costs, and

Table 2.1.3.-5. ANNUAL OPERATING COSTS FOR A 1 BILLION SCF EQUIVALENT STORAGE FACILITY FOR BELOWGROUND CONCRETE TANKS

1. Belowground Heaters, 300,000 kWhr/yr at \$0.01/kWhr	\$ 3,000
2. Maintenance, 0.5% of plant cost X \$2,800,000 plant cost	\$14,000
3. Operating Labor, 2/3 man X 4 X \$10,000	\$26,667
4. Operating Utilities, Electric Blowers Lights Pumps — 1,250,000 kWhr/yr at \$0.01/kWhr	\$12,500
5. Operating Supplies Lubricating oil Gland cooling water	\$ 1,000
6. Liquefaction Cost of Heat Leak Gas 365 X 10^6 CF at \$10,000/10^6 CF	\$25,550
Total	\$82,717

their appropriate depreciation, etc., are increased by a factor related to the increase in boil-off rate from 0.08 to 0.1%/day. Since an inert gas purge is not used with the insulation, the operating supplies' cost is reduced from \$5000 to \$1000.

The total annual operating cost for a storage tank of 1 billion SCF equivalent of LNG is \$609,717. Of this amount the most significant items are the depreciation, interest, taxes, insurance, etc., on the storage tank investment.

2.1.3.2.2. Frozen Pit

2.1.3.2.2.1. Description and Operating Characteristics

Cryogenic in-ground (CIG) storage has been the subject of a great deal of research and study in recent years. As a result of these efforts, this storage system has been brought to the point where accurate construction estimates can be made.

In 1960 Conch International Methane, Ltd., initiated a research program for the purpose of investigating this completely new concept in low-temperature storage: an excavated, earthen storage cavity with no insulation and no liner – in other words, a "bare hole in the ground with a cover on it" for containing liquid methane. Extensive research and theoretical analysis indicated the feasibility and suggested that the frozen earth would provide a completely adequate barrier for containing the −258°F liquid. After a preliminary small-scale test with ethylene at Shell Research Center in England, Conch decided that to confirm the theoretical data and to learn and obtain actual construction experience, a prototype installation should be attempted. Since the site chosen, Lake Charles, La., was in the heart of southern Louisiana's marshland, the project was very descriptively nicknamed "Operation Mudpie." It proved a success.

An 850-bbl container was constructed in soil with a roof constructed of metal. Figure 2.1.3.-8 shows the isothermal temperature gradients. The liquid temperature is −260°F.

In 1964, after the completion of this project, construction of a full-scale 240,000-bbl tank was initiated in Arzew, Algeria. The hole was 122 feet in diameter and 114 feet deep. The bottom was covered with concrete blocks over a layer of Foamglas insulation. Two submerged pumps, which discharge into a loading line, were located in the center of the floor. The following description is typical of the construction features for CIG tanks.

The first step is to accurately and completely determine the soil conditions as to the type of soil and the water table. There is still considerable question in the minds of many experts as to the necessary requirements. The recent failures of the CIG storage systems of Transcontinenal Gas Pipe Line Corp., at Carlstadt, N.J., and Tennessee Gas Pipeline Co., at Hopkinton, Mass., to meet their designed boil-off specifications have caused concern. The CIG systems operating at Arzew and at Canvey Island,

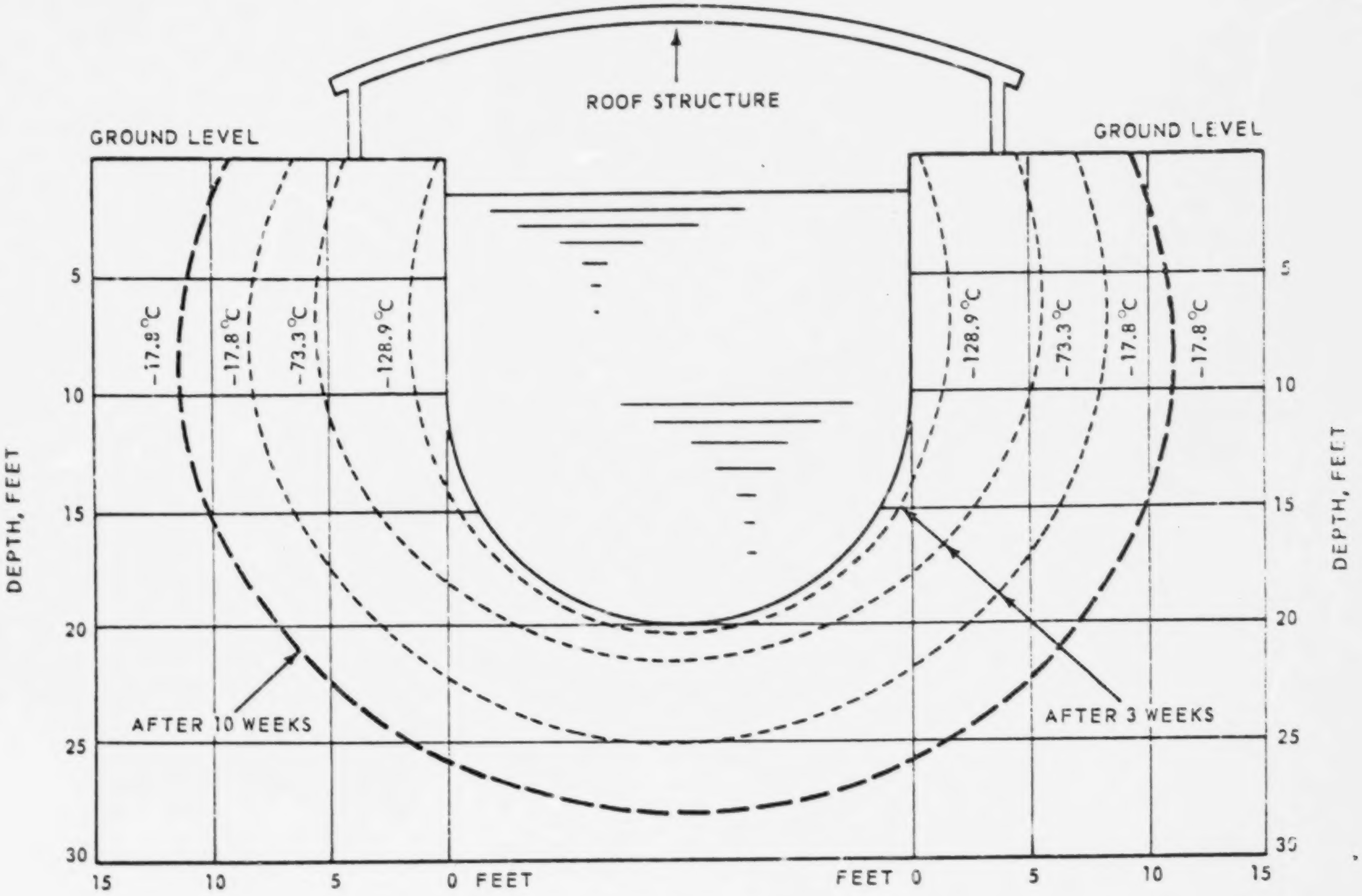

ROOF STRUCTURE
GROUND LEVEL
GROUND LEVEL
-17.8°C
-17.8°C
-73.3°C
-128.9°C
-128.9°C
-73.3°C
-17.8°C
-17.8°C
AFTER 10 WEEKS
AFTER 3 WEEKS
DEPTH, FEET
DEPTH, FEET
5
10
15
20
25
30
15
10
5
0 FEET
FEET 0
5
10
15

Figure 2.1.3.-8. GROUND TEMPERATURES AROUND LAKE CHARGES CAVITY[23]

England, have performed satisfactorily. The methods of construction in both of these cases were essentially the same. The major differences between the satisfactory and unsatisfactory installation have been the soil conditions, the water table conditions, and the depth of the prefreezing. The exact cause for the excessively high boil-off rate for the two CIG "failures" is not yet known.

The second step in CIG construction is the prefreezing of the container wall. The compressive strength of frozen earth can be as high as 3 times that of concrete. Freeze pipes are sunk into the soil to form a ring around the unfrozen core, which is then excavated. The depth of freezing is an important consideration. At Arzew the prefreeze refrigerant temperature was −13°F and the freeze pipes extended below the bottom of the tank.[34]

Figure 2.1.3.-9 illustrates the Arzew construction as licensed by Conch International Methane, Ltd. The water content of the soil was approximately 15-20% of the dry weight.

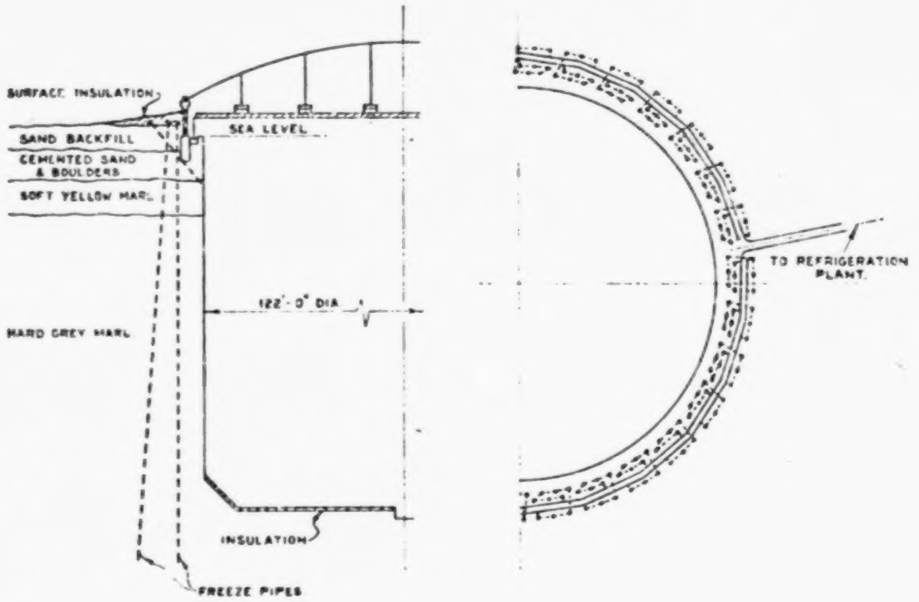

SURFACE INSULATION
SAND BACKFILL
SEA LEVEL
CEMENTED SAND & BOULDERS
SOFT YELLOW MARL
HARD GREY MARL
122'-0" DIA
TO REFRIGERATION PLANT
INSULATION
FREEZE PIPES

Figure 2.1.3.-9. GROUND FREEZING INSTALLATION[33]

The ground freezing installation consisted of 3 concentric rings, each containing 90 freeze tubes. The inner ring of tubes was placed on a diameter of 132 feet and was vertical. The second ring of tubes, on a diameter of 135 feet, was inclined outward at an angle of 4 degrees. These two rings of tubes went down to a depth of 131 feet.

In the third ring, 16-foot-long tubes were placed on a diameter of 147 feet and were inclined 40 degrees from the vertical inward. They were used to provide extra freezing in the shoulder, which had to support the weight of the roof structure during the latter stages of construction. The tubes were connected in series in groups of nine, each group including three pipes from each ring, to the inlet and outlet freeze header through isolating valves. The calcium chloride brine refrigerant was circulated through the headers at the rate of 900 gpm.

Approximately 6 months after the start of refrigeration, the refrigeration load had fallen off from an initial 275 tons to approximately 60 tons. The ice wall at the surface was then approximately 16 feet thick; at the bottom it was twice that. When the hole was sealed by installing the roof, the temperature inside the hole dropped to 0°F, a 13°F temperature differential between the freeze pipes and the air in the hole. Before LNG was put into storage, the temperature was further reduced to –76°F by passing cold gas from the liquefaction plant into the storage container. The LNG was then sprayed in at a slow rate compatible with the vapor handling capacity of the outlet system. After this, LNG was pumped in at a rate of approximately 1000 gpm.

Figure 2.1.3.-10 shows the position of the freezing isotherm, approximately 28°F, at different times after filling, which started in September 1965.[23]

The cooldown boil-off of a CIG hole is higher than that of an aboveground metal double-walled tank, even though the ground has been frozen before LNG was pumped into storage. Some authorities[45] believe that two important characteristics of CIG sites are –

1. Ground moisture must be sufficiently high so that, when the ground is frozen, the side walls will maintain a vapor tight seal. This generally means that the permanent groundwater level must be at or near the ground surface.

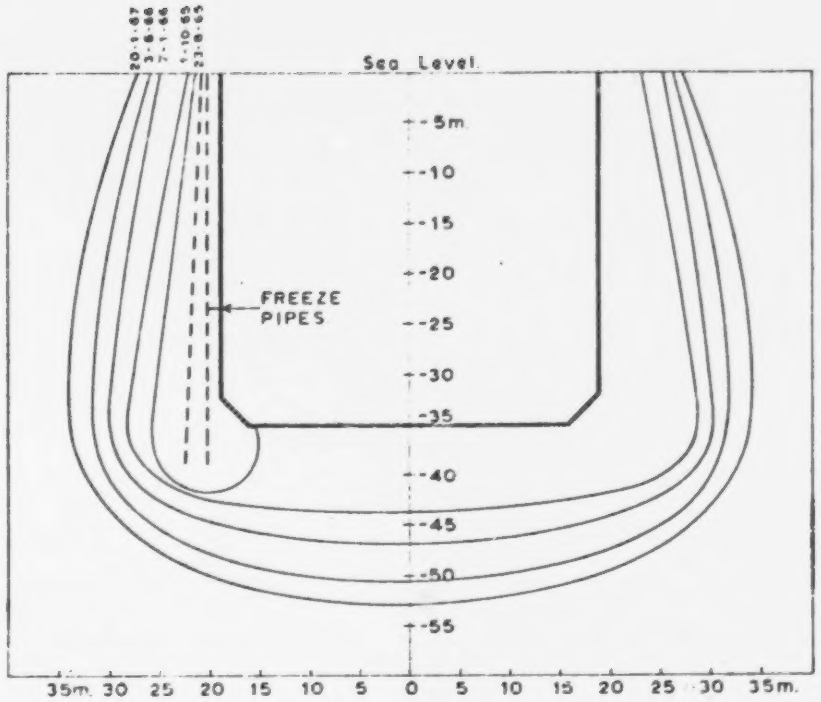


Figure 2.1.3.-10. POSITION OF FREEZING ISOTHERM AROUND GROUND STORAGE AT VARIOUS TIMES[33]

2. The bottom of the hole must be sealed by a relatively impervious earth strata, which will prevent water inflow at the bottom of the hole as a result of the large hydrostatic head.

Many other important factors are to be considered in the determination of the construction technique. Some analyses indicate the desirability of freezing even the core of the hole and excavating frozen earth. Although this is a more costly procedure, it may have some advantage in prestressing the initial frozen ring. This method is being used now on the four CIG storage systems being installed on Canvey Island in England, each of which has a capacity of 300,000 bbl. Figure 2.1.3.-11 illustrates the area of the frozen earth. Since the pipes installed in the center are 160-170 feet long and insulated to a depth of 130 feet, the cold energy is released only into the bottom or plug of the CIG system. A plan view of the 41.5-in.-OD pipes is shown in Figure 2.1.3.-12.

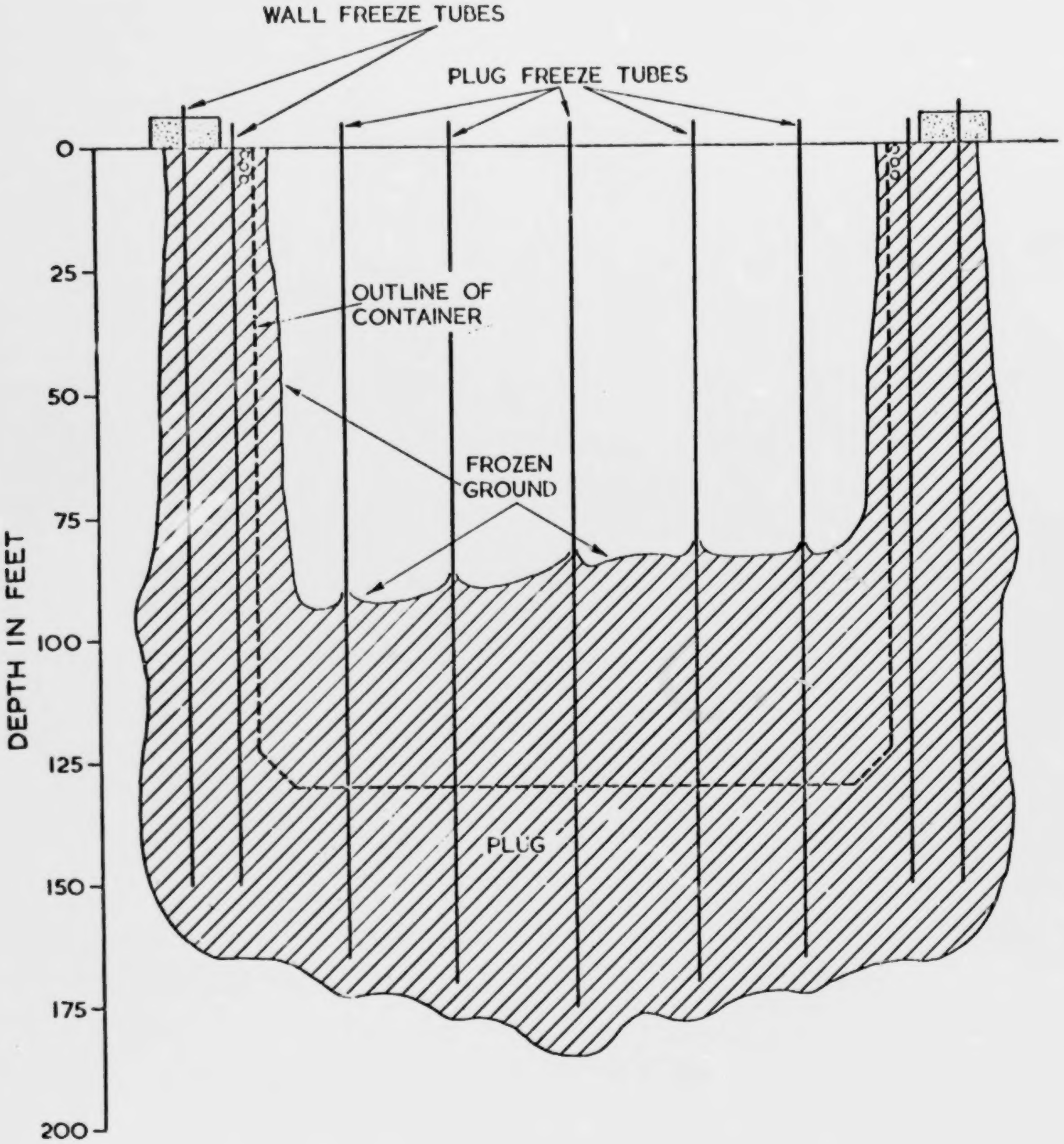

WALL FREEZE TUBES
PLUG FREEZE TUBES
OUTLINE OF CONTAINER
FROZEN GROUND
FROZEN GROUND
PLUG
DEPTH IN FEET
0
25
50
75
100
125
150
175
200

Figure 2.1.3-11. FROZEN-GROUND PROFILE AT START OF EXCAVATION[73]

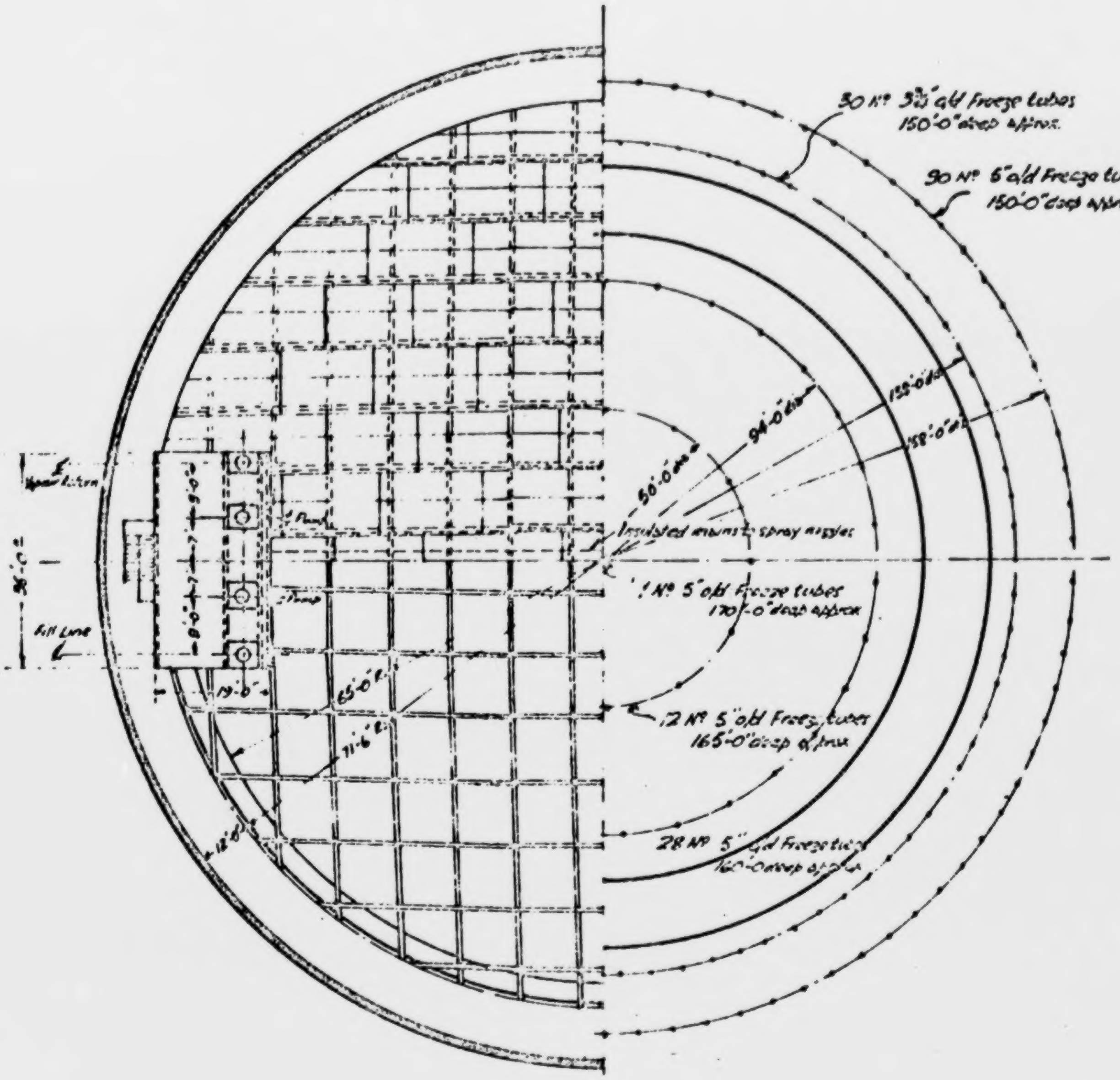

30 No 3½" o/d Freeze tubes
150'-0" deep approx.
90 No 6" o/d Freeze tubes
150'-0" deep approx.
133'-0" dia
138'-0" dia
94'-0" dia
50'-0" dia
Insulated mains to spray nozzles
1 No 5" o/d Freeze tubes
170'-0" deep approx
12 No 5" o/d Freeze tubes
165'-0" deep approx
28 No 5" o/d Freeze tubes
160'-0" deep approx
Fill Line
36'-0"
19'-6"
65'-0"
71'-6"
12'-0"
Pump
Pump

Figure 2.1.3.-12. PLAN VIEW OF 41.5-in.-OD PIPES OF LNG STORAGE SYSTEM ON CANVEY ISLAND 16

Other factors are the thermal conductivity of rocks. Table 2.1.3.-6 gives some test values. The results indicate little variation with temperature.

Table 2.1.3.-6. THERMAL CONDUCTIVITY OF ROCKS[70]

Type of Rock	Porosity, %	+75°F	−22°F	−112°F	−310°F
		— thermal conductivity, Btu-ft / sq ft-hr-°F—			
White chalk	30	--	1.09	0.93	1.04
Marl	3	0.97	1.03	0.94	0.96
Granite	1	1.05	1.25	0.95	0.79
Limestone	1	1.27	1.21	1.12	1.24

The Transco freeze pipes extended approximately 45 feet into the earth. They were not extended into the bedrock. Figure 2.1.3.-13. cross-section view of the Transco CIG storage system.

The high thermal conductivities of these rocks in contrast to foam insulation do not significantly affect the performance of the storage system, especially after the first year when a steady-state condition was achieved. The earth acts as a very thick insulant; thus boil-off rates comparable to aboveground (0.2%) are achieved.

The compressive strength of the frozen soil is needed to counteract the hydrostatic head of outside water, the head of LNG, the initial earth loads, and thermal stresses due to differential temperatures. The relatively high strengths of various soils are illustrated in Table 2.1.3.-7.

Table 2.1.3.-7. COMPRESSIVE STRENGTH OF ROCKS[70]

		Dry		Water Saturated	
Type of Rock	Porosity, %	+73°F	−292°F	+73°F	−292°F
		compressive strength, psi			
Cement	15	3,500	4,500	1,950	1,870
Sandstone	19	6,250	9,200	980	9,500
Limestone	3	23,500	40,000	18,700	27,200
"Gabbro"	6	14,750	21,500	11,600	23,000
Chalk	24	2,340	3,470	1,080	10,700

It should be emphasized that these are homogenous samples. They do not represent in situ results and thus could be considerably lower due to the presence of discontinuities, fissures, etc. The opening of natural faults would markedly affect the performance of a CIG storage tank.

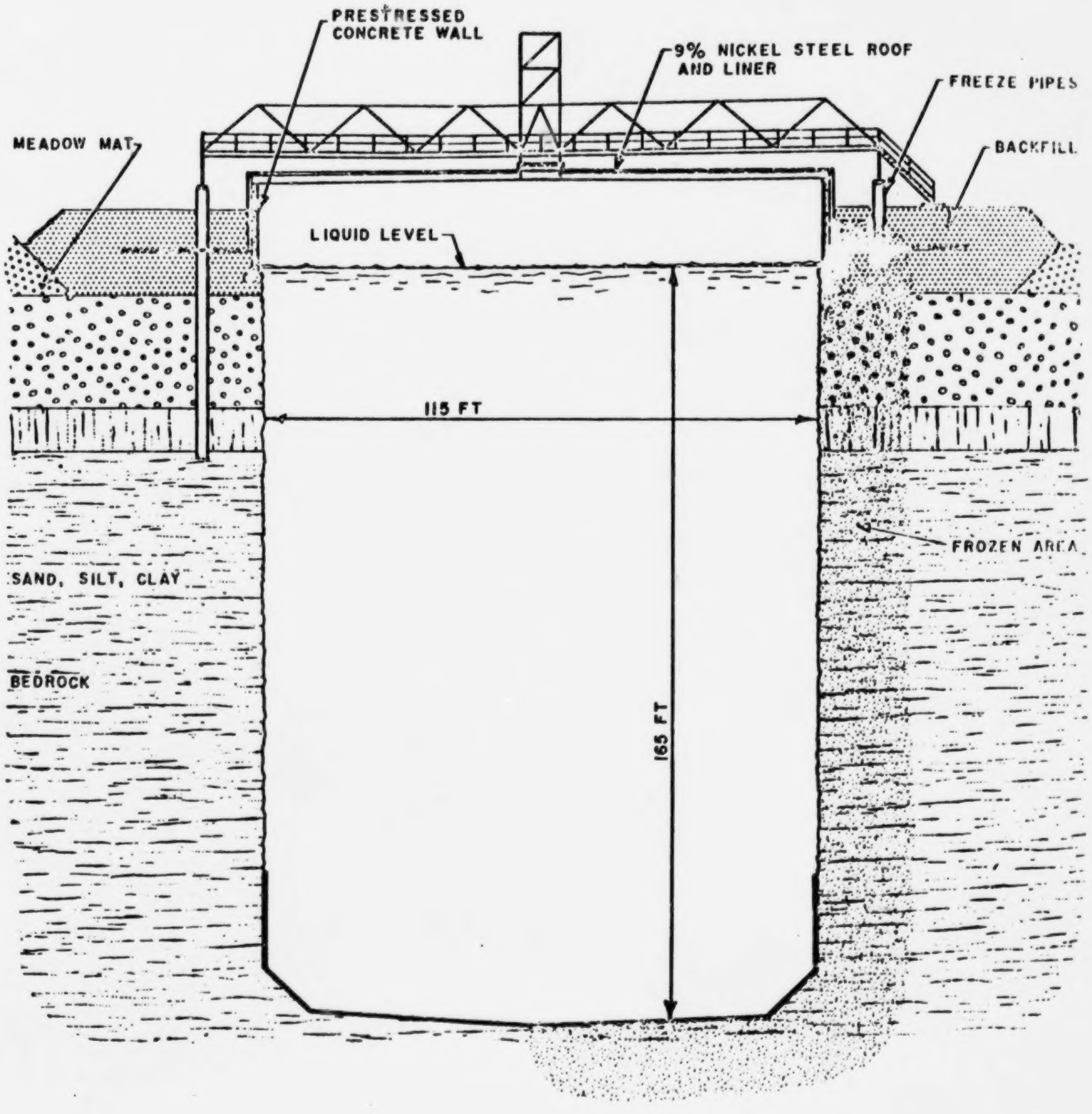

PRESTRESSED CONCRETE WALL
9% NICKEL STEEL ROOF AND LINER
FREEZE PIPES
MEADOW MAT
BACKFILL
LIQUID LEVEL
115 FT
165 FT
FROZEN AREA
SAND, SILT, CLAY
BEDROCK

Figure 2.13-43. CIG STORAGE UNIT FOR TRANSCO[43]

While the proper construction of the frozen hole is of primary importance, the aboveground portion, which supports the roof and its insulation, is also a factor. A reinforced concrete ring is used. It surrounds the hole and carries the roof structure. Reference 34 describes some of the details of roof construction at Arzew. The maximum internal design pressure is 15 in. of water. The roof material may be either 9% nickel steel or aluminum on the inner lining. Figure 2.1.3.-13 shows the roof construction and the use of backfill around the hole. Foamglas insulation is used on the roof. Leakage between the roof and the frozen ground is prevented by a stainless steel membrane.

After completion of construction, the storage tank is pressure-tested with air, the tank is purged with cold inert gas, and the inert gas is displaced by admission of LNG at the bottom and vented to the atmosphere. The ground surface external to the concrete wall is insulated and paved so there may be light traffic near the tank.

In spite of the difficulties of CIG storage in certain areas, this method still attracts considerable interest because of the safety and the potentially possible low unit cost per bbl.

2.1.3.2.2.2. Investment Costs

The investment cost of a frozen pit cryogenic in-ground (CIG) storage vessel can vary considerably for a given size depending on the geology at the site. Since engineering expertise is still being acquired and since there have been three vessels which have not performed as anticipated, actual cost data have not been released. To date, CIG storage systems built for peakshaving operations have not met their anticipated evaporation rates. When used for base-load operations (Algeria and United Kingdom), the system has, however, apparently been successful. One area of potential concern, however, has been the differential movements between the frozen pit and the unfrozen surrounding material. These movements may necessitate replacement or modifications of the storage piping.

An IGT cost estimate made in 1968 placed the cost of the vessel at \$9.00/bbl for 1 billion SCF of LNG equivalent storage capacity. This cost of \$2,565,000 is about 8% less than an aboveground metal vessel of the same capacity. The cost for the frozen pit includes the freezing system, holding refrigeration freezing costs, sitework, foundation and construction, roof, insulation, site improvements, transfer lines, instrumentation safety equipment, engineering services, contingencies, and applicable overhead. The appropriate portions of general plant facilities, warehouse, fences, roads, etc., are also charged against the storage facility.

It was assumed in this case that the freeze pipes were driven below the base of the vessel and that water influx through the base or walls of the excavation did not occur. The evaporation rate was assumed at 0.32% of the storage capacity per day.

This cost figure must be used with caution due to the great uncertainties in evaluation of individual items which vary with geologic conditions at the site.

The cost of a CIG frozen pit is only 8% less than an aboveground metal tank for a 285,000-bbl capacity; however, for larger sizes, such as the 1,200,000-bbl pit for Philadelphia Gas Works, the anticipated savings should be significantly greater.

2.1.3.2.2.3. Operating Costs

The annual operating costs for a frozen pit (CIG) storage system on the same cost basis as is used for the aboveground metal tank, Section 2.1.3.1.1.3., is given in Table 2.1.3.-8. The total annual operating cost is $246,000. The total annual operating cost is about 3 times that of an aboveground metal tank of equal storage volume.

Table 2.1.3.-8. ANNUAL OPERATING COSTS FOR A FROZEN PIT (CIG) STORAGE FACILITY

1. Maintenance		$12,825
2. Operating Labor		$26,667
3. Operating Utilities, Electric		
Blowers Lights Pumps	4,000,000 kWhr/yr at $0.01/kWhr	$40,000
4. Operating Supplies		
Lubricating oil Gland Cooling water, etc.		$ 3,000
5. Liquefaction Cost of Heat Leak Gas, 2336 X 10^6 CF at $70.00/$10^6$ CF		$163,520
	Total	$246,012

2.1.3.2.3. Cavern

2.1.3.2.3.1. Description and Operating Characteristics

Two different cavern system concepts have been developed:

a. The IGT atmospheric pressure storage utilized primarily for peakshaving operations, and

b. The Gaz de France pressure storage for base-load operations.

Both systems, as in the case of frozen pits, are highly dependent on geologic conditions at the site for establishing both technical feasibility and economics.

The IGT concept[38] is based on the approach that since LNG can be stored at atmospheric pressure, there is no need to mine a deep cavern as is done in liquefied petroleum gas (LPG) storage. Mining techniques for large-volume near-surface operations favor a large, incline entry to a room-and-pillar opening as being the most economic. Insulation may be used to reduce the boil-off rates to values obtained with other insulated systems.

The most economical method of entry per unit of length is through a horizontal tunnel. If the depth is not excessive, an inclined shaft entry can be utilized. The size of the entry is governed by the size of the loading and transportation equipment, and may typically be 13 feet wide by 17 feet high. Once the entry is dug to the desired depth, the actual excavation of the cavern may progress in stages. A typical excavation using a room-and-pillar mining system is shown in Figure 2.1.3.-14. The progress of the excavation in three stages is depicted in Figure 2.1.3.-15. The mining cycle includes drilling, blasting, scaling, and loading. It is estimated that an average daily excavation of 550 cubic yards can be readily maintained, thus permitting rapid and economical schedules. The excavating cost per storage bbl using the above technique is significantly lower than that for a vertical-shaft pressure cavern.

The economic advantages of cavern storage of LNG are contingent on the successful utilization of a liner-insulation system that would contain the LNG and reduce the large heat-influx rates from bare rock.

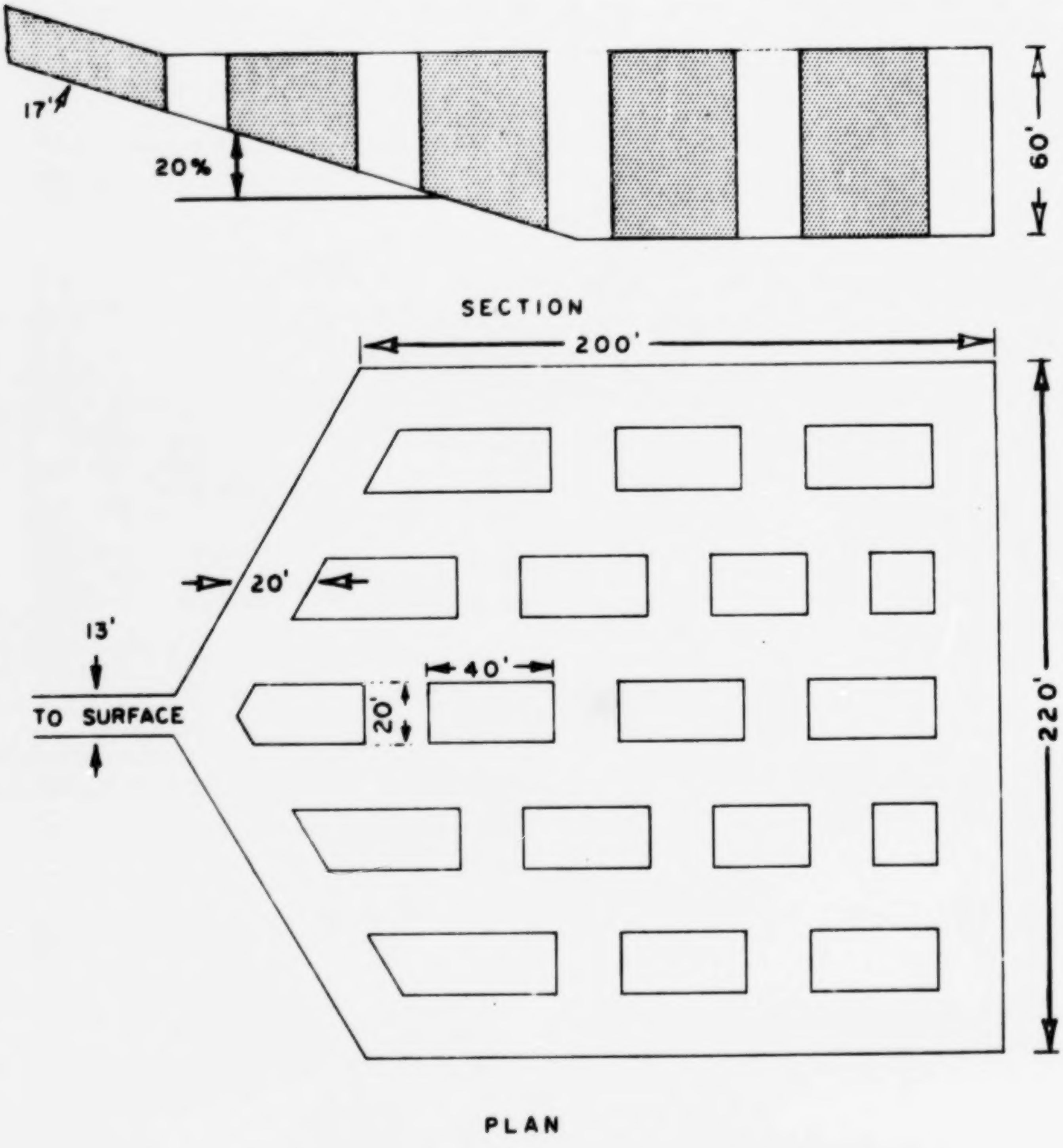
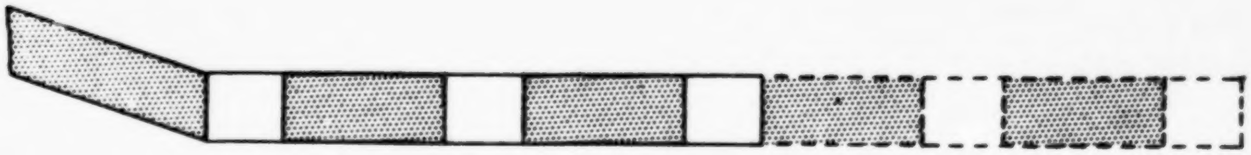
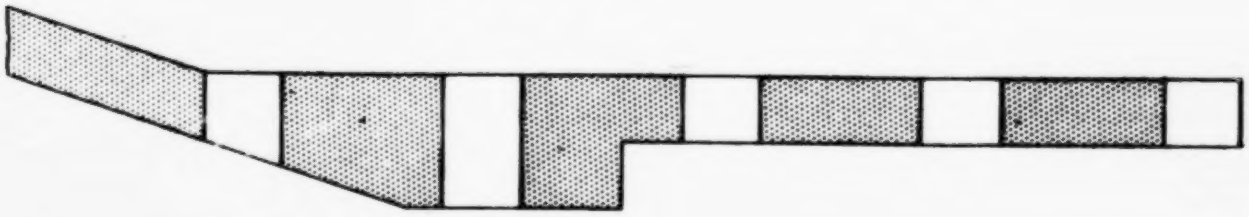

17'
20%
60'
SECTION
200'
220'
20'
13'
TO SURFACE
40'
20'
PLAN

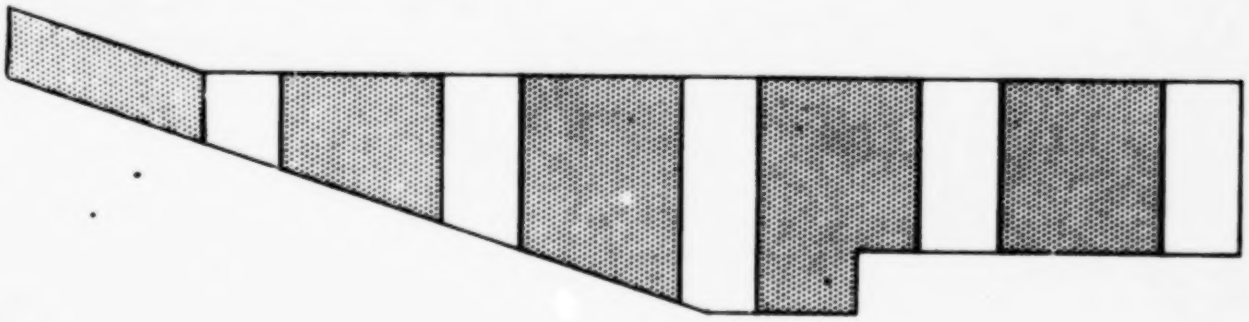
Figure 2.1.3.-14. TYPICAL EXCAVATION USING ROOM-AND-PILLAR MINE SYSTEM



STAGE I TOP LEVEL ADVANCEMENT



STAGE 2 SECOND LEVEL ADVANCEMENT



STAGE 3 THIRD LEVEL ADVANCEMENT

Figure 2.1.3.-15. EXCAVATION PROGRESS

The entire system consists of the load-bearing cavern wall surface, a water vapor barrier, an adhesive (if required) to bond the insulation to the barrier, foamed insulation, and a flexible, impermeable (LNG barrier) liner fastened to the foamed insulation.

The water barrier must provide a seal for fractures in the rock surface which might be up to 1/4 in. wide which is impermeable to water and water vapor under a hydrostatic head up to 90 psi. It must be able to bond to a wet surface, and remain integral to 32°F, at which point frozen pore water forms an adequate moisture barrier.

The insulation must have a low thermal conductivity, resist spalling or cracking due to the cryogenic temperatures, and be easily and economically applicable. A polyurethane foam insulation which has a low thermal conductivity ($k = 0.14$ Btu-in./hr-sq ft-°F) is suitable based on its physical and chemical properties.

Combination plastic-metal laminates will best meet the containment problem. Such laminates composed of Mylar, aluminum, and Dacron have high ultimate tensile and tear strength, high elongation, and a very low permeability to methane gas. The bonding of the liner is achieved by a simple heat seal, using a tape composed of the same material as the liner but with a special cryogenic adhesive.

During the construction of a pilot cavern for demonstration purposes, it was found that the water barrier would not perform satisfactorily under the site conditions. An alternative system design to deal with this problem was developed. A highly permeable cemented aggregate was placed between the rock wall and the insulation supports. Sumps and pump discharges to handle water influx were provided. Adequate insulation is placed between the stored liquid and the permeable water-handling zone to prevent freezing. This system, although expensive provides a demonstrated solution to water influx. Where water influx is not encountered, the unmodified and less expensive design may be used.

The Gaz de France cavern is designed to provide large-volume storage for base-load operations at the point of offloading.

The storage concept[4] is shown in Figure 2. 1. 3.-16. It consists of a vertical shift (1) which is used as an access for excavating the storage galleries (2). The shaft is concrete-lined, and is provided with a sump at its base to collect water that seeps in from higher water-bearing strata. The storage galleries are excavated in impermeable strata.

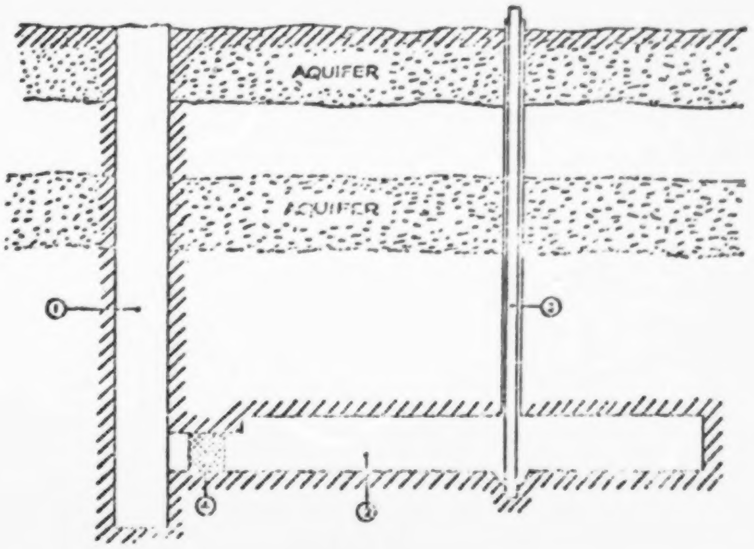


Figure 2. 1. 3.-16. GAZ de FRANCE CONCEPT FOR COMMERCIAL CAVERN STORAGE [4]

The profile of the galleries depends on the structural nature of the rock, while the spacing and arrangement is determined from heat transfer considerations.

The operation of the storage cavern is carried out through casements (3). These are from the ground surface and consist of two concentric pipes; the inner one is used for filling and withdrawal, while the annular space between the two serves for injection or release of gas as required for pressure equilibrium. A lock with manhole (4) is constructed at the storage gallery entrance.

A short time after filling the gallery with LNG, water is allowed to freeze slowly at the base of the shaft. After a plug of ice is formed, the vertical shaft is filled with water. This procedure equalizes the pressure on either side of the lock. The gallery is operated at a pressure corresponding to the hydrostatic pressure of the overburden.

Gaz de France concluded that for Le Havre (the depot which will accept LNG from refrigerated tankers), underground storage would be the most advantageous method. However, this conclusion became definitive only after the storage methods for the depot at Le Havre had already been decided upon. Although cavern storage at Le Havre is not an immediate necessity, Gaz de France is prepared, if the need arises, to proceed with such a project.

Gaz de France has outlined some of the procedures and principles anticipated for putting a cavern into operation:

1. The storage galleries will be dried by circulating dry air; wall dryness will be checked by measurement of temperature and relative humidity of the air entering and leaving the reservoir.
2. If insulation, such as foamed plastic, is to be installed on the wall, this will be done after the first drying; a second drying will follow.
3. The manhole at the lock on the shaft end of the reservoir will be closed and the cavity flushed with dry nitrogen, ultimately leaving it slightly pressurized; the shaft is flushed with carbon dioxide.
4. Introduction of LNG into the storage galleries then initiates the cooling.
5. When the shaft side of the lock is below the freezing point, the shaft will be gradually filled with water so as to form an ice plug at the bottom of the shaft; the shaft is then completely filled with water.

The Le Havre cavern will operate at a pressure sufficient to restrict, and perhaps even eliminate, evaporation caused by heat supplied by the surrounding rock. This will allow siphoning of the liquid at any time. It is further planned to tap the reservoir only by siphoning liquid because drawing-off evaporated gas may lead to heavy product concentration in the liquid, and possibly to sedimentation of solids on the bottom.

2. 1. 3. 2. 3. 2. Investment Costs

On the same basis as aboveground metal tanks, Section 2. 1. 3. 1. 1. 2., the investment cost for an IGT cavern, exclusive of land and cost of money, is estimated by IGT at approximately $1, 800, 000, providing a storage capacity of 1 billion SCF of LNG equivalent. This cavern has an evaporation rate of 0. 05%/day and is constructed in an impermeable strata thereby requiring no water handling. If water handling and hence the alternative design were required, which is the most probable situation, the cost would be $3, 215, 000.

No investment costs have been made available for the Gaz de France storage concept. Based on mining costs for LPG caverns, the investment cost should be approximately $2, 900, 000 for the 1 billion SCF of LNG equivalent.

2. 1. 3. 2. 3. 3. Operating Costs

No operating cost estimates are available for the cavern systems. Exclusive of the charge for interest, depreciation, taxes, etc., the operating costs for the IGT cavern should approximate those for an aboveground metal tank where no water handling is required. Where water handling is required, the additional charge for power consumption will vary with each site. The operating costs for a Gaz de France-type of cavern, exclusive of charges for interest, etc., will approximate those for a frozen pit (CIG) system.

2.1.3.2.4. Metal Tank — In-Ground Storage

2.1.3.2.4.1. Description and Operating Characteristics

The basic elements of the belowground double-walled tank will be identical to those of an aboveground tank: the inner shell will be of suitable material for cryogenic service, the outer shell will be of carbon steel, and the annular space will be filled with insulating material. A ring-wall-type foundation with or without piling will be used. Heating coils will be installed to prevent freezing of frost-susceptible soils. External fittings will pass the roof of the tank, rather than the wall.

Construction of a double-shell steel vessel either belowgrade or partially belowgrade with a berm would require alteration of the design of the outer shell. The outer shell will have to withstand the additional loading due to the backfill and/or berm. The load will be several times greater than the external load on aboveground tanks. The outer shell will also have to be protected against corrosion. Another required modification will be the substitution of submersible pumps such as those used in belowground concrete tankage.

The evaporation rate from such a vessel would be equivalent to an aboveground tank, and its operating procedures and characteristics will be similar to those of a belowground concrete vessel.

2.1.3.2.4.2. Investment Costs

The investment cost for a belowground steel tank may be considered as consisting of the normal or aboveground costs plus the additional increment for placing the tank belowground. Diking costs must then be subtracted from the cost because a dike will no longer be required. The cost of the excavation, backfill, and/or berm will depend upon the precise configuration of the system. The tank might be constructed completely belowgrade with selected backfill. In this case the excavation costs would be maximized.

The outer tank cost would also be a maximum as its structural load would be the highest. If the tank were built ongrade with a berm that was not placed against the tank, the excavation and outer shell costs would be minimal and berming cost maximum. Depending upon the soil and groundwater conditions at the site and the cost of and need for selected backfill, an economic optimum can be obtained. As this optimum is dependent on the site, no investment costs are presented.

2.1.3.2.4.3. Operating Costs

No operating costs have been prepared for the in-ground double-walled steel tanks. Exclusive of depreciation, interest, insurance, taxes, etc., the operating costs would be comparable to an aboveground double-walled metal tank. This is based on a tank which is above the water table where no water-handling problems exist. As the investment costs are, however, substantially higher than for an aboveground tank, the overall operating costs will reflect this additional cost.

2.1.4. LNG Transportation Systems

After natural gas is converted to a liquid by cooling, it is readily transportable as a cryogenic liquid. The transportation equipment must meet the requirements imposed by the chemical composition and physical state of the liquid but beyond that it poses no problems other than those found in transporting any other liquid fuel.

2.1.4.1. LNG Ocean Tankers

The vital link in the transportation route to supply the necessary large volumes of LNG from foreign sources are the deepwater marine tankers. Although cryogenic tankers have been in service for a number of years, primarily in the ammonia or LPG trade, the history of LNG tankers dates only from the first tests of the Methane Pioneer, conducted in the late 1950's. The Pioneer, now renamed the Aristotle, was rebuilt in 1958 for methane service and made several trips from the United States Gulf Coast to the United Kingdom.

LNG tankers are unique among deepwater tankers in several respects. They are rated to operate at lower temperatures than any other liquid-gas carriers. The cargo space is heavily insulated, reducing the effective cargo volume. The low-density cargo necessitates special ballasting systems. Tankers presently in use, or under construction, use a conventional steam turbine propulsion system, but the fuel system is usually modified to allow the use of boil-off natural gas from the cargo tanks.

The first LNG tankers in service, other than the experimental Methane Pioneer, were of modest size. The Methane Progress and Methane Princess had a capacity of approximately 27,400 cubic meters, or 172,000 bbl. The capacity is steadily increasing until the latest ships to be construction by the Phillips-Marathon group for the Alaska-Japan project have a capacity of 450,000 bbl. The size, capacity, registry, etc., of the tankers now in LNG service and those under construction for the service are listed in Table 2.1.4.-1. Figure 2.1.4.-1 gives a comparison of the various LNG tankers with an oil supertanker.

Table 2.1.4.-1. LNG TANKERS UNDER CONSTRUCTION OR IN SERVICE

Ship Name or Hull Number	Service	Owner	Manager	Flags of Registry	Cargo Capacity: Liquid, cu m	Cargo Capacity: Barrels	Cargo Capacity: Liquid, 1000 SCF	Dimensions: LOA	Dimensions: Breadth	Dimensions: Draft
Aristotle (Methane Pioneer)	Charter	Antarctic Gas Inc.	Liquid Carriers, Ltd.	Panamanian	5,123	32,223	180.9	338 ft 9 in.	50 ft 4 in.	23 ft 6-1/2 in.
Jules Verne	Arzew, Algeria, to Le Havre, France	Gaz Marine. S. A.	Gaz S. A.	French	25,500	160,390	900.5	659 ft 6 in.	81 ft 7 in.	24 ft 8-1/8 in.
Methane Princess	Arzew, Algeria, to Canvey, U. K.	Conch Methane Tankers, Ltd.	Shell Tankers (U. K.), Ltd.	British	27,400	172,341	967.6	621 ft 0 in.	81 ft 8 in.	26 ft 0 in.
Methane Progress	Arzew, Algeria, to Canvey, U. K.	Methane Tanker Finance, Ltd.	Shell Tankers (U. K.), Ltd.	British	27,400	172,341	967.6	621 ft 0 in.	81 ft 8 in.	26 ft 0 in.
Pythagore	--	Gaz S. A.	--	French	630	3,963	22.2	191 ft 3 in.	28 ft 0 in.	11 ft 6 in.
Hull No. 1645 Hull No. 1646 Hull No. 1947	Marsa el Brega, Libya, to LaSpezia, Italy, and Barcelona, Spain	Prora Transporti. S. p. A.	--	Italian	40,000	251,592	1,412.6	681 ft 6 in.	96 ft 0 in.	28 ft 6 in.
Hull No. 211	Marsa el Brega, Libya, to Barcelona, Spain	Nav. de Productos Licuados S. A.	--	Spanish	40,000	251,592	1,412.6	674 ft 8 in.	96 ft 0 in.	28 ft 6 in.
Hull No. 516 Hull No. 517	Kenai, Alaska, to Negishi, Japan	Polar LNG Shipping Co.	--	Liberian	71,500	449,720	2,525.0	800 ft 0 in.	111 ft 6 in.	32 ft 9 in.

Table 2. 1. 4. -1, Cont. LNG TANKERS UNDER CONSTRUCTION OR IN SERVICE

Ship Name or Hull Number	Tanks		Unloading Pumps			Year Built	Propulsion		Speed, nautical mph	Fuel, tons/day
	Number	Type	Number	Type	Capacity		Type	Rating, shaft horse-power		
Aristotle (Methane Pioneer)	5	Insulated, rectangular	5 2	Deepwell booster	26,417 gal/hr at 85 ft liquid	1945 Rebuilt 1958	2-S. A. 6-cylinder, Nordberg	2,000 brake horsepower	9.5	9 Gas oil
Jules Verne	7	Cylindrical, vertical, insulated	14	--	--	1965	2 Steam turbines, Parsons	11,500 brake horsepower	17.5	85 Fuel oil
Methane Princess	9	Rectangular, aluminum, insulated	9	Submerged	486,072 gal/hr at 269 ft	1964	2 Steam turbines	12,500	17.25	75 Fuel oil
Methane Progress	9	Rectangular, aluminum, insulated	9	Submerged	--	1964	2 Steam turbines	12,500 brake horsepower	17.25	75 Fuel oil
Pythagore	2	Membrane, refrigerated, rectangular	2	--	--	1964	Two 4-S. A. each 6-cylinder, M. A. N.	1,200 brake horsepower	11	4 Gas oil
Hull No. 1645 Hull No. 1646	4	Rectangular, insulated	8	Independent	1,056,680 gal/hr at 394 ft	1969	Steam turbine	15,000	18	90
Hull No. 1947	4	Rectangular, insulated	8	Independent	1,056,680 gal/hr at 394 ft	1970	Steam turbine	15,000	18	90
Hull No. 211	4	Rectangular, insulated, aluminum	8	Deepwell	1,268,016 gal/hr at 427 ft	1969	Steam turbine	15,000	18	--
Hull No. 516 Hull No. 517	6	Insulated, membrane	-	--	--	1969	Steam turbine	20,000	17	Fuel oil as complement to evaporated gas

Source: Liquid Gas Carrier Registrar, London: H. Clarkson, 1968.

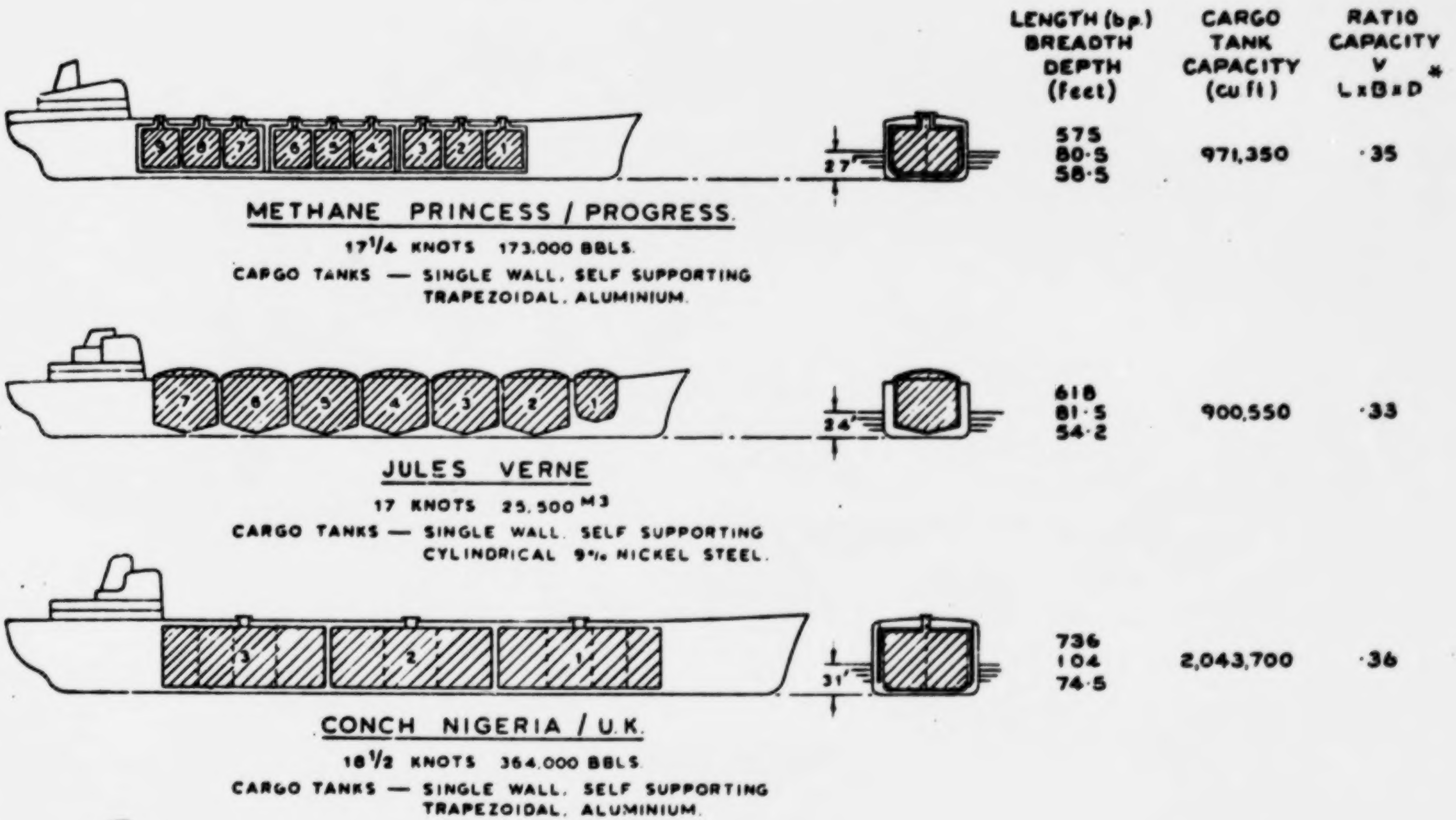

LENGTH (b.p.) BREADTH DEPTH (Feet)
CARGO TANK CAPACITY (cu ft)
RATIO CAPACITY V / L x B x D *
27'
575
80·5
58·5
971,350
·35
METHANE PRINCESS / PROGRESS.
17 1/4 KNOTS 173,000 BBLS.
CARGO TANKS — SINGLE WALL. SELF SUPPORTING TRAPEZOIDAL. ALUMINIUM.
24'
618
81·5
54·2
900,550
·33
JULES VERNE
17 KNOTS 25,500 M3
CARGO TANKS — SINGLE WALL. SELF SUPPORTING CYLINDRICAL 9% NICKEL STEEL.
31'
736
104
74·5
2,043,700
·36
CONCH NIGERIA / U.K.
18 1/2 KNOTS 364,000 BBLS.
CARGO TANKS — SINGLE WALL. SELF SUPPORTING TRAPEZOIDAL. ALUMINIUM.
32'
700
100
68·5
2,043,700
·425
CONCH MEMBRANE DESIGN.
18 1/2 KNOTS 364,000 BBLS.
CARGO TANKS — SINGLE WALL MEMBRANE, INVAR OR STAINLESS STEEL.
27·5'
640
96
60·5
1,403,650
·375
ESSO LNG. TANKERS.
17 1/2 KNOTS 250,000 BBLS.
CARGO TANKS — SELF SUPPORTING TRAPEZOIDAL. ALUMINIUM.
43'
781
106
59·5
2,000,000
·65
TYPICAL 53,500 T.D.W. OIL TANKER.
16 KNOTS
CARGO TANKS — INTEGRAL WITH SHIPS STRUCTURE.

Figure 2. 1. 4. -1. COMPARISON OF OCEAN TANKERS[19]

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* Ratio Capacity, $V = \frac{\text{Cargo Tank Capacity}}{\text{Length X Breadth X Depth}}$

2.1.4.1.1. Freestanding Tanks

2.1.4.1.1.1. Description

The first LNG tankers built utilized the so-called free-standing tank construction. That is, conventional self-supporting insulated tanks were prefabricated and mounted in the hull of the ship. The Methane Progress, typical of this class of construction, carries nine separate aluminum cargo tanks. These tanks were built in the prefabrication shop, tested prior to mounting in the hull, and finally lifted into place. The insulation installed between the cargo tank wall and the steel bulkhead is comprised of balsa wood panels sandwiched between layers of fiber glass. Details of the hull construction, tanks, and insulation system are shown in Figure 2.1.4.-2.

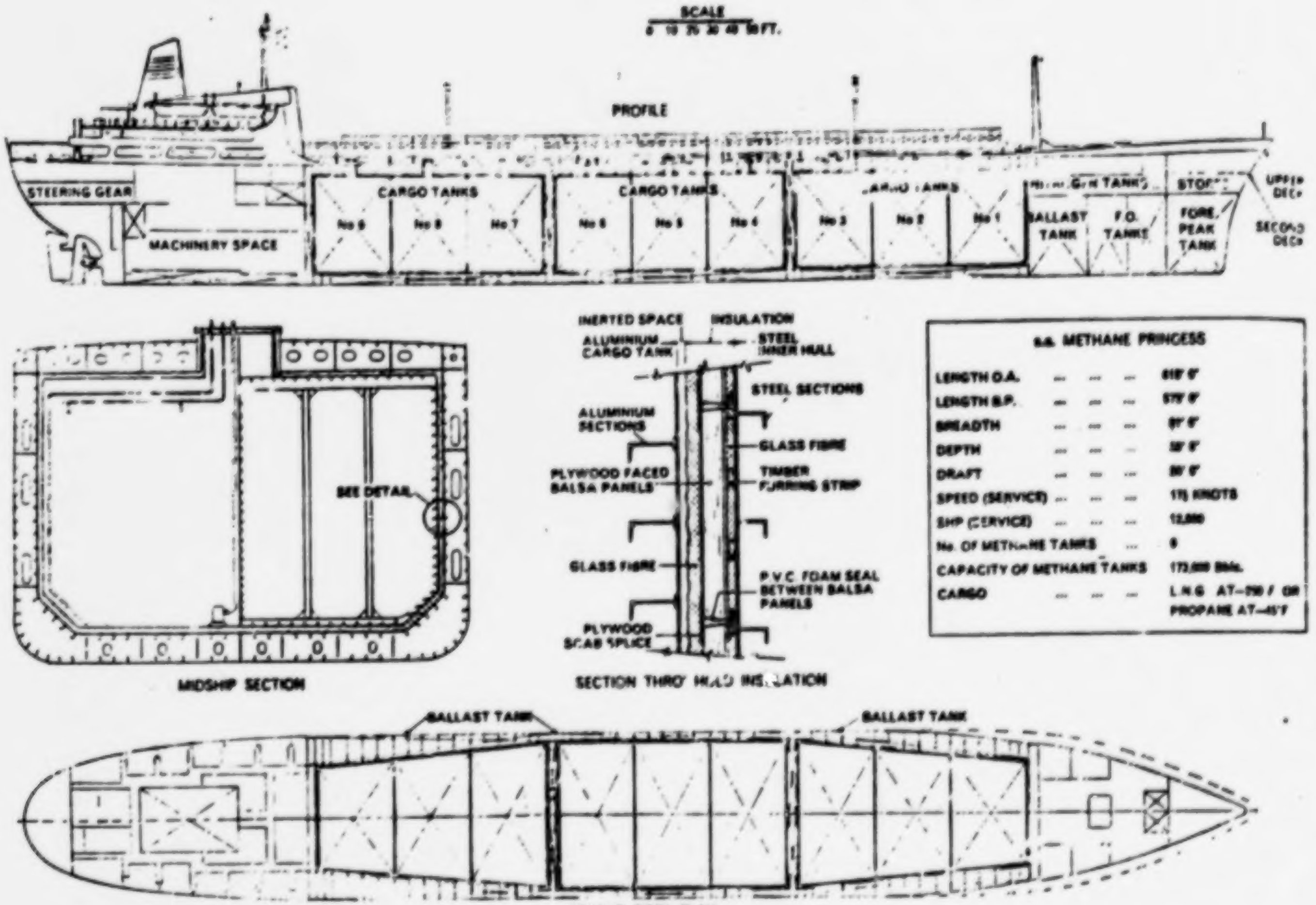

SCALE
0 10 20 30 40 50 FT.
PROFILE
STEERING GEAR
MACHINERY SPACE
CARGO TANKS
No 9
No 8
No 7
CARGO TANKS
No 6
No 5
No 4
CARGO TANKS
No 3
No 2
No 1
BALLAST TANK
F.O. TANKS
STORES
FORE PEAK TANK
UPPER DECK
SECOND DECK
SEE DETAIL
MIDSHIP SECTION
INERTED SPACE
INSULATION
ALUMINIUM CARGO TANK
STEEL INNER HULL
STEEL SECTIONS
ALUMINIUM SECTIONS
GLASS FIBRE
PLYWOOD FACED BALSA PANELS
TIMBER FURRING STRIP
GLASS FIBRE
P.V.C. FOAM SEAL BETWEEN BALSA PANELS
PLYWOOD SCAB SPLICE
SECTION THRO' HEAD INSULATION
s.s. METHANE PRINCESS
LENGTH O.A. — — — 618' 6"
LENGTH B.P. — — — 579' 6"
BREADTH — — — 81' 6"
DEPTH — — — 58' 6"
DRAFT — — — 26' 6"
SPEED (SERVICE) — — — 17½ KNOTS
SHP (SERVICE) — — — 12,000
No. OF METHANE TANKS — 9
CAPACITY OF METHANE TANKS 172,000 Bbls.
CARGO — — — L.N.G. AT -258°F OR PROPANE AT -45°F
BALLAST TANK
BALLAST TANK
SECOND DECK PLAN

Figure 2.1.4.-2. GENERAL ARRANGEMENT OF ALGERIA-U.K. LNG CARRIERS[32]

The free-standing tank construction combined with the low density of LNG requires that LNG tankers have larger hulls with low draft and high freeboards. Since the cargo tanks cannot be used to carry water ballast, as oil tankers do, separate ballast tanks must be provided at some cost and space sacrifice.

The low temperature requires special materials, usually aluminum or 9% nickel steel, for the cargo tanks, and special arrangements must be made to accommodate the large expansion and contraction induced by thermal cycling of the tanks. The tanks, as well as all other equipment on board, must meet not only the containment and cryogenic requirements but the shipboard environment. They must be able to withstand continuous vibration, sustained pitching and rolling— up to 35% from the vertical— and dynamic forces as great as 1.6 G. They are subjected to stresses induced by the hull deflection, constantly attacked by the corrosive seawater, and suffer from the lack of readily available competent and trained repairmen.

2.1.4.1.1.2. Investment Costs

Tanker investment or construction costs can vary significantly depending upon the availability of ship ways at the time the order is placed and the location of the shipyard. If the ships must be built in United States' yards, in order to comply with the Jones Act requirement for interstate shipment, costs will be higher than if the tankers are built in foreign yards. It has been estimated that U.S. construction costs can be as much as 40% higher.

Tankers cost also vary significantly with capacity. A single propulsion train, with only slight incremental cost increases, can serve a wide range of tanker sizes. Figure 2.1.4.-3 shows the variation in tanker cost with size when the free-standing tank design is used.

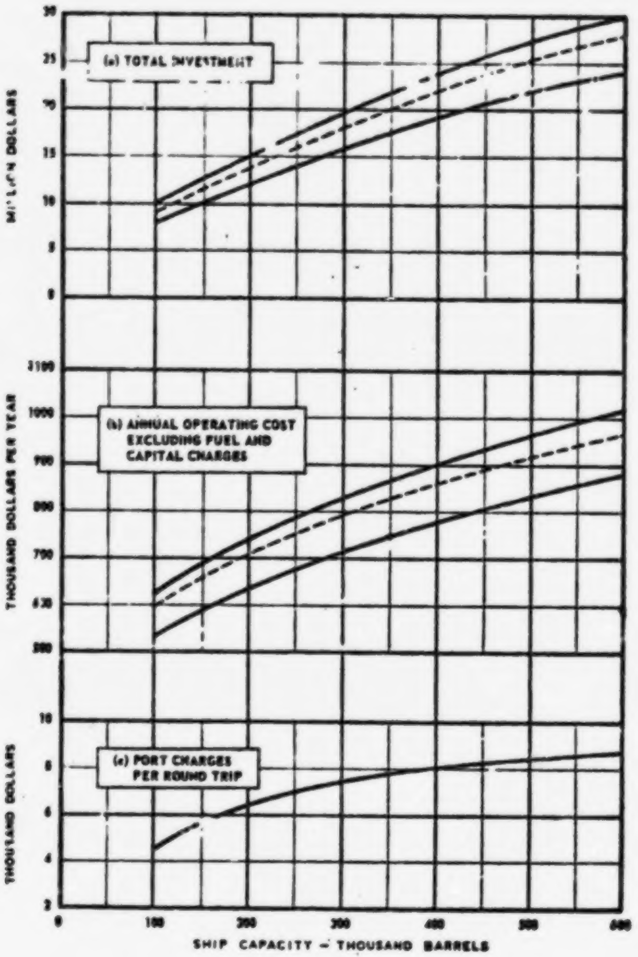


Figure 2. 1. 4. -3. LNG CARRIER COSTS [32]

2. 1. 4. 1. 1. 3. Operating Costs

A detailed analysis of operating costs for LNG tankers is given in Section 3. 2. 4. 2. 2. The major items of operating expense, regardless of the tanker design, are fuel, crew expenses, ship operating expenses, and port charges. For a large LNG tanker, about 600, 000 bbl, typical operating expenses are about $100, 000 /month when non-U. S. crews are used. Major items of expense are crews' wages, 15 %; insurance, 40 %; and repair and maintenance reserve, 15 %.

2.1.4.1.2. Membrane Tankers

The first LNG tankers built used freestanding cryogenic tanks. However, in an effort to reduce tanker costs and increase hull space utilization, several companies undertook development of a new storage tank concept in which the tank could be more nearly moulded to the contours of the ship's hull. Since the inner tank need not be self-supporting, it can be made of a very thin, although impermeable, metal barrier. Thus the term "membrane tanker" developed. This type of tank was fully tested in the Pythagore and is being installed in the two tankers which are under construction for the Alaska-Japan trade.

2.1.4.1.2.1. Description

Details of the membrane type tanks being installed in the Polar LNG Shipping Company tankers are shown in Figures 2.1.4.-4 and 2.1.4.-5. The load, the weight of the liquid, is transferred by the load-bearing insulation to the outer hull. The so-called membrane system consists of an inner membrane containment vessel made of Invar (36% nickel steel), perlite insulation, and an outer, wooden box that contains the perlite and transmits the cargo load to the hull. To conform to Coast Guard regulations a secondary barrier that is designed to contain the cargo in the event of failure of the primary barrier was installed.

2.1.4.1.2.2. Investment Costs

While an extensive cost history has not yet been developed for tankers of this type, it is anticipated that cost per unit of LNG shipping capacity will be somewhat lower than the free-standing tank construction. One intermediate size tanker now on order will cost approximately $70/bbl of capacity. This is estimated to be about 10% less than a freestanding tank tanker would cost.[74]

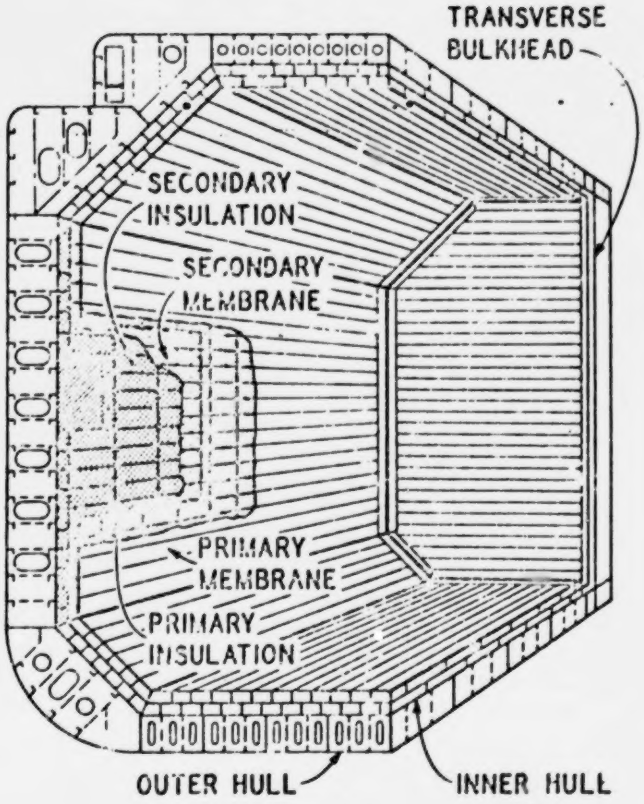

TRANSVERSE BULKHEAD
SECONDARY INSULATION
SECONDARY MEMBRANE
PRIMARY MEMBRANE
PRIMARY INSULATION
OUTER HULL
INNER HULL

Figure 2.1.4.-4. CUTAWAY OF A METHANE TANKER CARGO TANK [15]

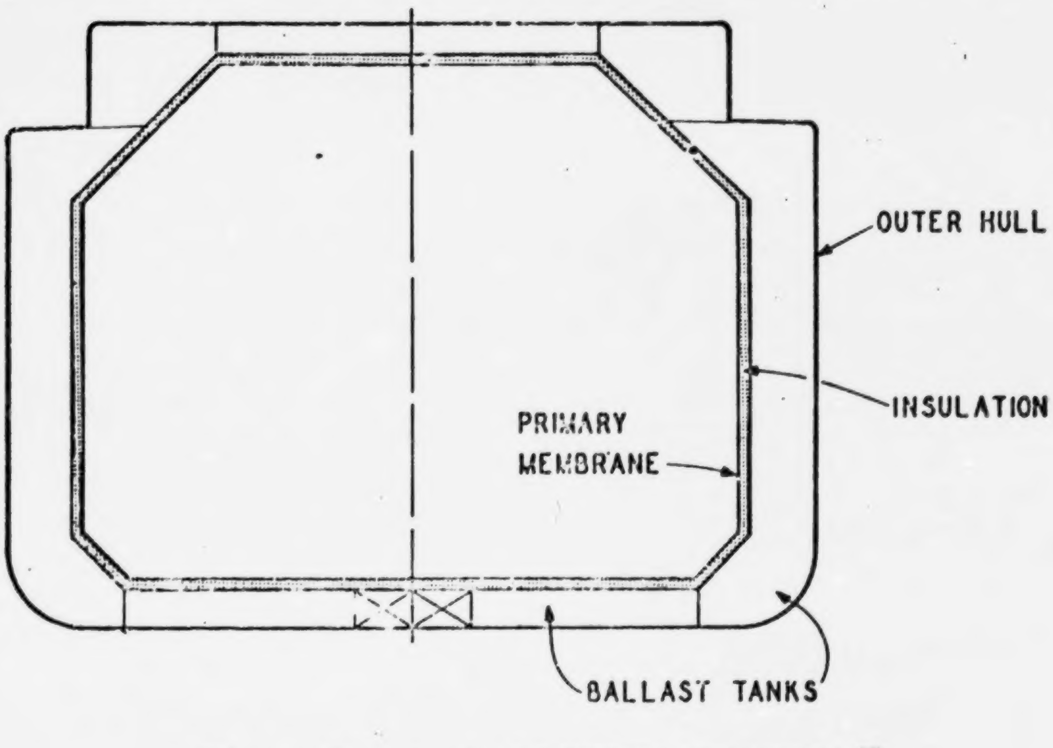

OUTER HULL
INSULATION
PRIMARY MEMBRANE
BALLAST TANKS

Figure 2.1.4.-5. MIDSHIP SECTION [15]

2.1.4.1.2.3. Operating Costs

Operating costs for membrane tankers are essentially the same as for any other type. The costs are detailed in Section 3.2.4.2.2.

2.1.4.1.3. Propulsion Systems

All of the LNG tankers, except the experimental Aristotle and Pythagore, use conventional steam-turbine-driven single-screw propulsion systems. The Polar LNG tankers will be equipped with bow thrusters to facilitate docking in Alaskan ports where tugboat service is not readily available. Gas turbines may offer an attractive alternative to steam turbines and several conceptual designs of LNG tankers using gas turbines that have been developed.

2.1.4.1.3.1. Steam Propulsion Systems

The typical steam propulsion system found on an LNG tanker hardly differs from that of an oil tanker. The engine room of the Methane Progress is typical and contains the single necessary change; the engine room atmosphere is continuously evacuated and monitored for methane gas.

The two dual-fuel boilers, Foster Wheeler Type-ESP II, produce 49,000 lb/hr of 600-psig 850°F steam under normal conditions and 73,500 lb/hr under maximum load. The furnaces are rated under conditions of simultaneous firing of oil and gas but either can be fired independently. These boilers, the first ever built for marine, gas-fired service, also have a feed gas heater to raise the temperature of the boil-off gas from −60° to +140°F.

Two steam turbines develop 12,500 shp at 107 rpm, and 13,750 shp at 110 rpm.

2.1.4.1.3.2. Gas Turbines

Gas turbine marine propulsion systems are of recent vintage. In the 1950's one cargo ship was converted to gas turbines, but it was not until the 1960's that a cargo vessel, the Admiral Callaghan, was designed and built specifically with gas turbine propulsion. The principal advantage is that a much larger power output, necessary if high speeds are to be achieved, is possible with gas turbines. Fuel efficiency is likely to be somewhat less than with steam turbines, but maintenance costs may be reduced.

It has been estimated[63] that substitution of a gas turbine propulsion system for steam turbines could save over $2 million. Machinery cost savings represent 43% of this, although the balance comes from decreased hull costs, saving on design costs, etc.

In addition, machine weight would be reduced by from 400 to 500 tons, engine room space by 25%, and the ship's volume overall reduced by about 10%.

2.1.4.2. LNG Trailers and Tanker Trucks

Liquefied natural gas can be transported overland either by pipeline, tank trucks, or tank cars. Tank trucks, designed for road transport of LNG, have been operated successfully in the Soviet Union for a number of years. These trucks are used to distribute LNG to areas far removed from pipelines and to fuel trucks that are designed to run on natural gas. In the United States, as a logical consequence of the development of tank trucks for cryogenic usage with liquid oxygen, nitrogen, and hydrogen, tank trucks have recently been constructed for the transporting of LNG. The first instance of such truck transport of LNG was undertaken by San Diego Gas & Electric Co. in 1966. The trailer-mounted cryogenic tank filled with LNG was used with a vaporizer to supply an uninterrupted flow of gas to a service area where the main transmission line had been shut off for maintenance work. Since then San Diego Gas & Electric has successfully transported, in conjunction with Cryogenic Enterprises, Ltd., three trailers of LNG from San Diego to Squamish Valley, Canada. Numerous other shipments of LNG, for special test purposes, have been made between points in the United States.

Regular LNG transportation service, by truck, is now being provided between the Chula Vista, Calif., LNG plant of San Diego Gas & Electric Co. and a trailer park 80 miles away in Borrego Springs, Calif. The same company also transports LNG by truck to a military base some distance from its LNG plant.

2.1.4.2.1. Description

LNG tanker trucks and trailers must meet the same criteria of safe containment of a flammable cryogenic liquid that ocean tankers must meet. Over-the-road trailers also must stay within stringent dimensional and weight limitations, but otherwise are quite similar in design concept to any other transportation equipment. An inner containment vessel made of cryogenic metal is surrounded by an insulating space. This may be a vacuum insulation or may be filled with a suitable insulating material such as perlite. An optimized trailer design which has been developed is a trailer having a capacity of 10,000 gal, equivalent to 918,000 CF of natural gas. The tank has an operating pressure of 30 psig. Other trailers, ranging in size from 6000 to 11,000 gal, have been designed.

The trailer can be fitted with a gravity-fed vaporizing coil so that a pressure transfer system can be used for unloading. In the event that the receiving vessel is at a higher pressure than the trailer tank operating pressure, a cryogenic liquid transfer pump mounted on the truck can be used.

2.1.4.2.2. Investment Costs

The investment costs involved with an LNG trailer include only the trailer itself. Standard typical trailer costs are —

Size, gal	Cost, $
6,000	25,000
6,000	31,000
11,000	50,000
11,625	37,200

2.1.4.2.3. Operating Cost

Operating costs for LNG truck trailer transportation include the cost of operation and maintenance of the tractor to haul the tanker, labor, and overhead charges for the personnel involved and capital charges on the trailer. Typical operating and maintenance costs for a trailer at the size required for an 11,000-gal trailer are $0.08/mile.

One estimate[17] places the cost of LNG transport by tanker truck at $0.15/1000 CF for 100-mile haul and $0.27/1000 CF for a 200-mile haul.

2.1.4.3. Railroad Tank Cars

Although LNG has not yet been transported by railroad tank car in the United States, tank car designs have been prepared.

2.1.4.3.1. Description and Operating Characteristics

A tank car designed for LNG service in accordance with the Department of Transportation Specification DOT 113C60W would carry 113,750 lb of LNG, equivalent to approximately 2.75 million CF. This is the largest sized tank car which can be used in unlimited interchange service. The internal stainless steel vessel is suspended within a carbon steel outer shell. The annular space is filled with perlite and evacuated. The inner tank suspension system is designed to minimize heat leak. The heat leak is so low that the total through the insulation, supports, and piping will result in a pressure rise of only 30 psi over a 40-day period.

2.1.4.3.2. Investment Costs

LNG railroad tank cars have been quoted at a price of $29,000 for a 34,200-gal capacity car. Since this is the largest car which can be used, there is little incentive to provide smaller cars that would be more costly per unit of capacity.

2.1.4.3.3. Operating Costs

Freight rates have not yet been established for LNG transportation. However, it appears that a Class 13 rate would be applicable to LNG. Rates are reduced by a "mileage rate allowance" that compensates the car owner for use of his car. For an LNG car, the allowance would be $0.18/loaded mile. Class 13 rates are approximately $0.35/100 lb for a 300-mile haul and $0.50/100 lb for a 600-mile haul.

In terms of energy transportation costs, the cost of an LNG railroad transport is estimated to be $0.23/1000 CF for a 200-mile haul and $0.34/1000 CF for a 600-mile haul.

2.1.4.4. Cryogenic Barges

Another mode of water transportation that has been considered, but not yet put into operation, is that of barging the LNG. Barges may be designed for either seagoing service or in land waterways, and while each has its own structural needs, stiffening members, etc., most design features are common to both. Barges have been designed that essentially consist of double-walled metal tanks with insulation in the space between the walls, similar to those built to transport the liquid hydrogen for NASA. These barges could be used as a means for distributing LNG from a large central liquefaction plant to smaller local storage facilities.

Cryogenic barges offer an attractive and economic extension of the waterborne LNG transportation system. Deepwater tankers can serve only those ports with suitable berthing facilities, but shallow-draft barges could easily receive LNG from the tanker and, at low cost, move the fuel inland to river or lake ports. Their use has been considered by a number of companies. The first plan of Constock was to move LNG from the Gulf Coast to Chicago via barge; at least one company has seriously considered using a seagoing barge. Although no LNG barges have as yet been built, the cryogenic technology in hand should allow for their design and construction when needed.

2.1.4.4.1. Description

Even though no barges have been built, a conceptual design has been developed. A barge capable of carrying LNG equivalent to 100 million CF (28,000 bbl) would be 293 ft long and 50 ft wide. The cargo tank, built of 9% nickel, would be 18 ft high, 40 ft wide, and 218 ft long. The outer carbon steel shell would be 22 ft high, 44 ft wide, and 222 ft long. The space between the two shells would be filled with a suitable thermal insulation.

Although barges are not subjected to all of the rigors of a deepwater tanker, they must, nevertheless, be designed to withstand the stresses induced by sudden movements, minor collisons, and rough handling.

2. 1. 4. 4. 2. Investment Costs

No definitive estimates have as yet been made for LNG barges. However, very rough estimates for an inland waterway service barge indicate costs may be in the range of \$30-\$50/bbl of LNG capacity. Seagoing barges could be expected to cost 25 to 40% more.

2. 1. 4. 4. 3. Operating Costs

LNG transportation by barge over inland waterways could be a common carrier operation, in which case the operating costs would be the filed tariff for the particular haul. However, the more likely condition is that the barges would be owned, or at least chartered, by the customer. If this is the case, operating costs would include tow boat charter fees, docking fees, and loading and unloading costs. No estimates have been made of what these costs might be. However, characteristically barge transport costs are only a fraction of road or rail transport cost. If the combination of navigable water access at both terminals and a manageable volume throughput coincide, barges could be economically attractive.

2.1.5. Vaporization Systems

It has been traditional to vaporize LNG before burning it. Whether this will be necessary or desirable for future LNG-fueled power plants is open to question. Existing LNG base-load plants serve gas pipelines and, therefore, vaporization of LNG is mandatory. However, short LNG pipelines between a tank and a power plant boiler may provide some simplification of equipment, though this would require special burners. Fortunately, there are a variety of reliable vaporizers available and these do not represent an inordinate part of the invested capital of the LNG system. Fired vaporizers absorb approximately 2% of the Btu that they transmit to the pipeline, so that the operating cost is not large. The primary requirements for a regasification system are —

1. LNG pumps
2. Heat exchangers
3. Controls
4. Odorants

Among the systems discussed, the pumps, controls, and odorants are quite similar. The main differences are in the methods of heat exchange.

2.1.5.1. Fired Vaporizers

2.1.5.1.1. Direct-Fired Vaporizers

2.1.5.1.1.1. Description

Figure 2.1.5.-1 illustrates an LNG vaporization system. The pump is typically a centrifugal or turbine pump rather than a rotary. Multistaging is used to obtain the desired pressure; provisions are made to ensure adequate NPSH. Conventional cryogenic service pumps are used. The typical peakshaving vaporization system is provided with 50 to 100% standby because of the relatively low cost of providing this additional performance guarantee. However, this is not the case for base-load plants.

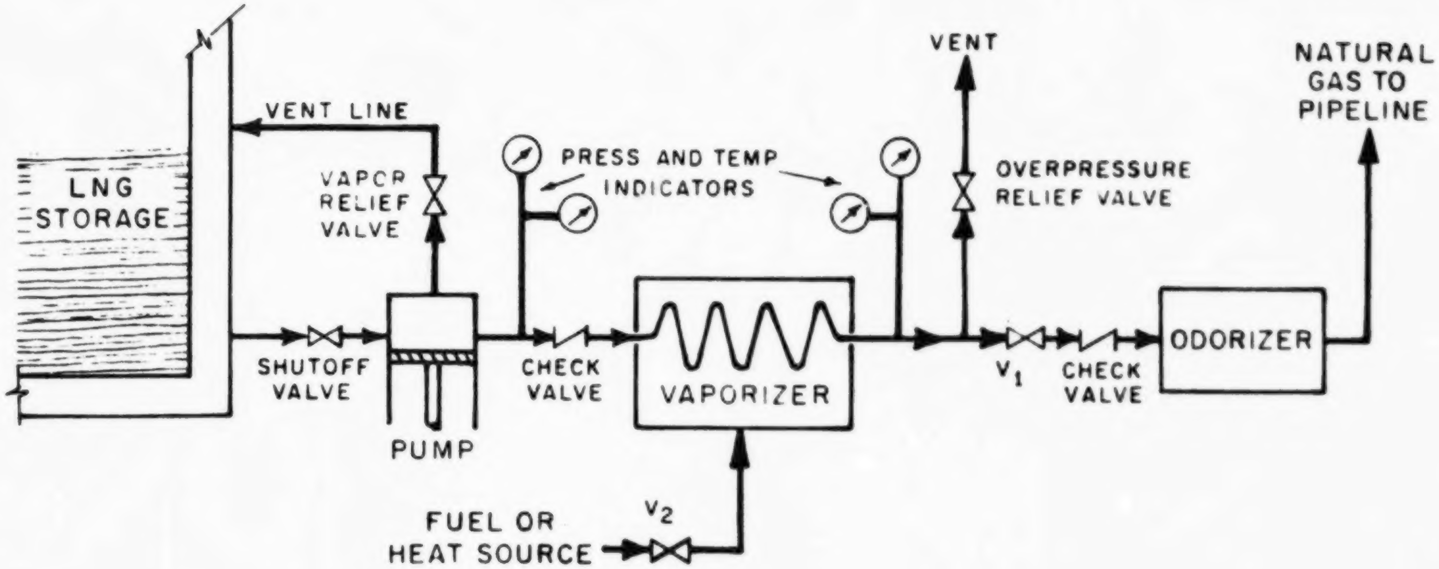

LNG STORAGE
VENT LINE
VAPOR RELIEF VALVE
SHUTOFF VALVE
PUMP
PRESS AND TEMP INDICATORS
CHECK VALVE
VAPORIZER
FUEL OR HEAT SOURCE
V_2
VENT
OVERPRESSURE RELIEF VALVE
V_1
CHECK VALVE
ODORIZER
NATURAL GAS TO PIPELINE

Figure 2.1.5.-1. SIMPLIFIED FLOW DIAGRAM OF TYPICAL LNG VAPORIZER SYSTEM[44]

Figure 2. 1. 5. -2 shows a typical installation and Figure 2. 1. 5. -3 illustrates the principles of the direct-fired heat exchanger. Stainless steel finned tubing is stacked in a rectangular chamber. LNG comes in at the bottom and the gas leaves at the top tubes. The combustion of air and gas is completed in the conical combustion chamber and stoichiometric air/fuel rates are used. The hot inert gas at velocities up to 400 ft/s is discharged into the tube bank. A large mass of hot gas at about 1000°F flows in a dual elliptical path and maintains a heat flux of approximately 20,000 Btu/hr/sq ft of pipe surface. The gas circulates an average of 5 times around this bank before going out the stack.

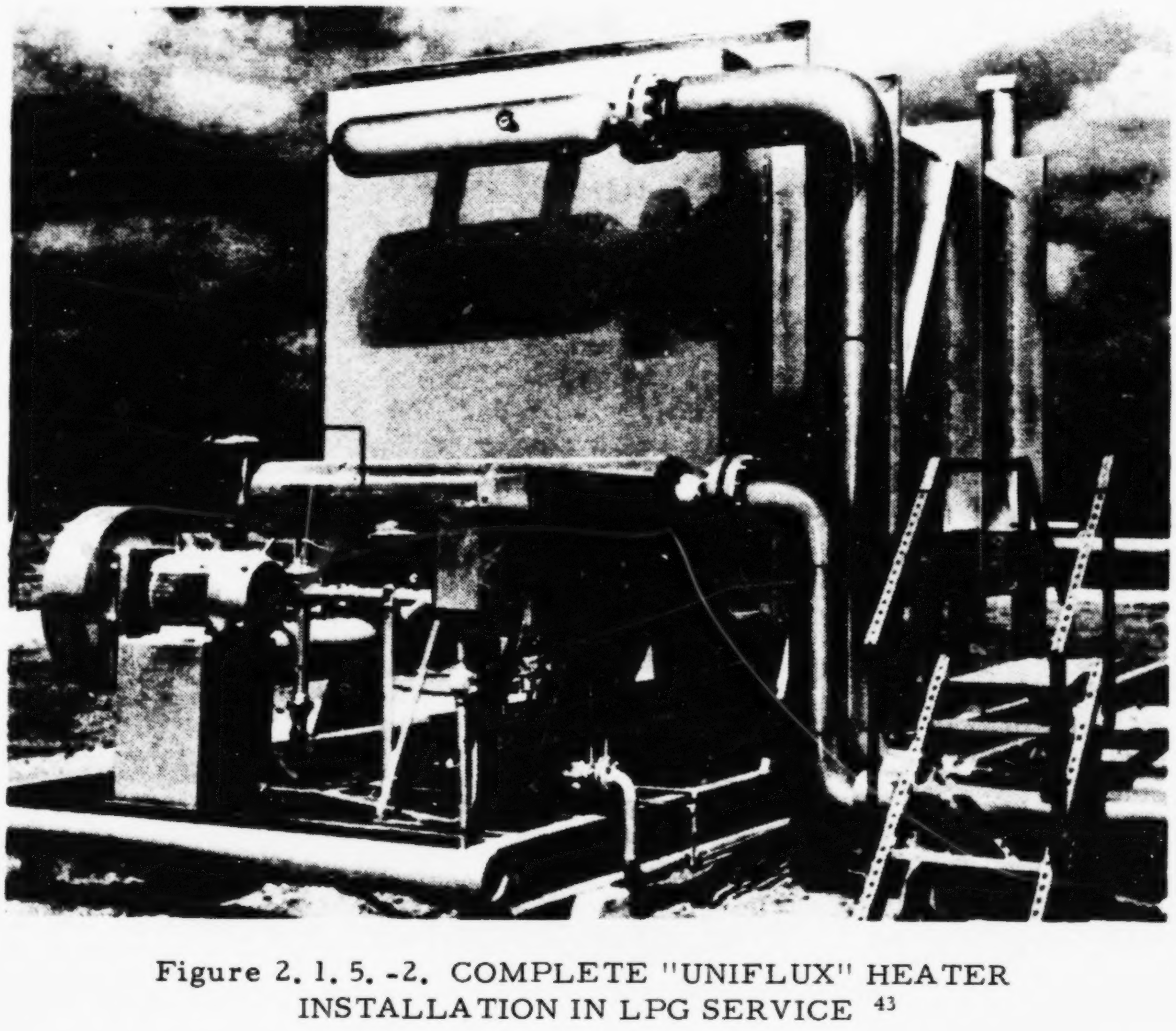


Figure 2. 1. 5. -2. COMPLETE "UNIFLUX" HEATER INSTALLATION IN LPG SERVICE [43]

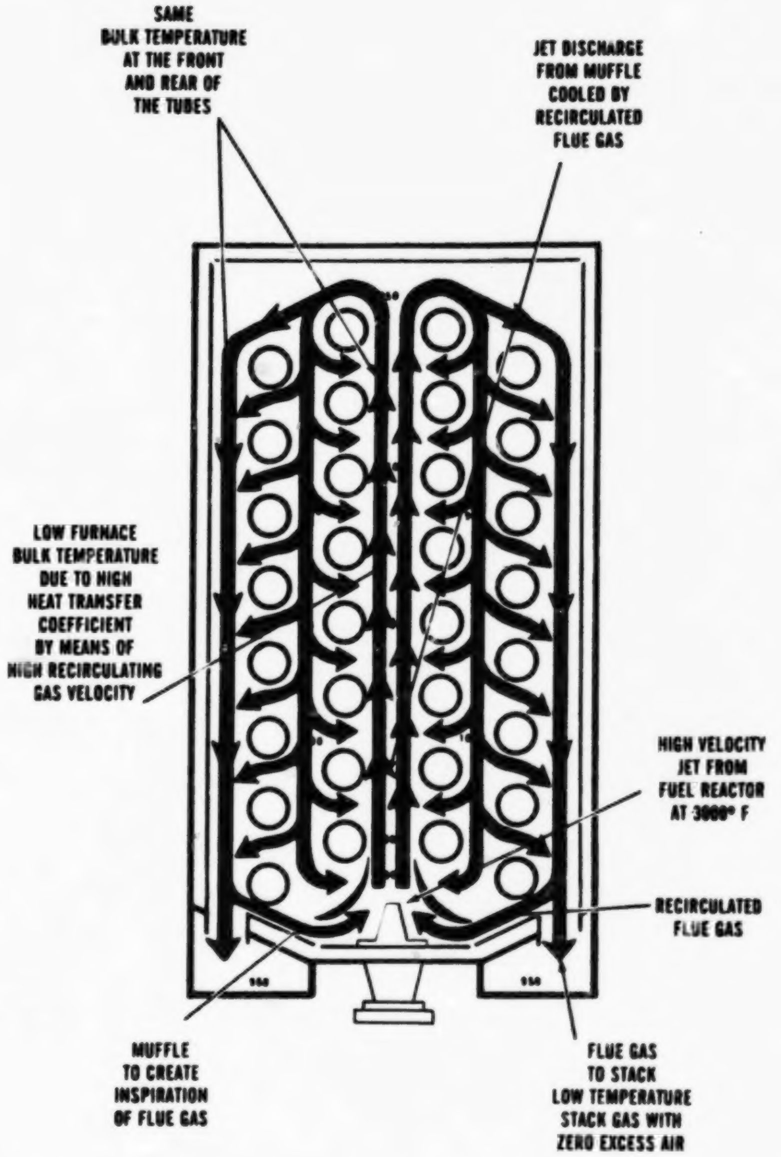

SAME BULK TEMPERATURE AT THE FRONT AND REAR OF THE TUBES
JET DISCHARGE FROM MUFFLE COOLED BY RECIRCULATED FLUE GAS
LOW FURNACE BULK TEMPERATURE DUE TO HIGH HEAT TRANSFER COEFFICIENT BY MEANS OF HIGH RECIRCULATING GAS VELOCITY
HIGH VELOCITY JET FROM FUEL REACTOR AT 3000° F
RECIRCULATED FLUE GAS
MUFFLE TO CREATE INSPIRATION OF FLUE GAS
FLUE GAS TO STACK LOW TEMPERATURE STACK GAS WITH ZERO EXCESS AIR

Figure 2. 1. 5. -3. DIAGRAMMATIC CROSS SECTION OF DIRECT-FIRED CONVECTIVE HEAT EXCHANGER (Black, Sivalls & Bryson "Uniflux" Design) [2]

The large temperature difference and the high rate of inert gas scrubbing combine to make this type of unit the most compact type of vaporizer. The small mass of the system is a factor tending to improve controllability. Ice formation on the fins and tubing is avoided by the maintenance of a stoichiometric fuel/air ratio which also ensures maximum gas temperature. In the typical application (300-psia discharge pressure of vaporized LNG) the heat input is divided into three approximately equal sections: 1) LNG liquid warming, 2) vaporization, and 3) superheat.

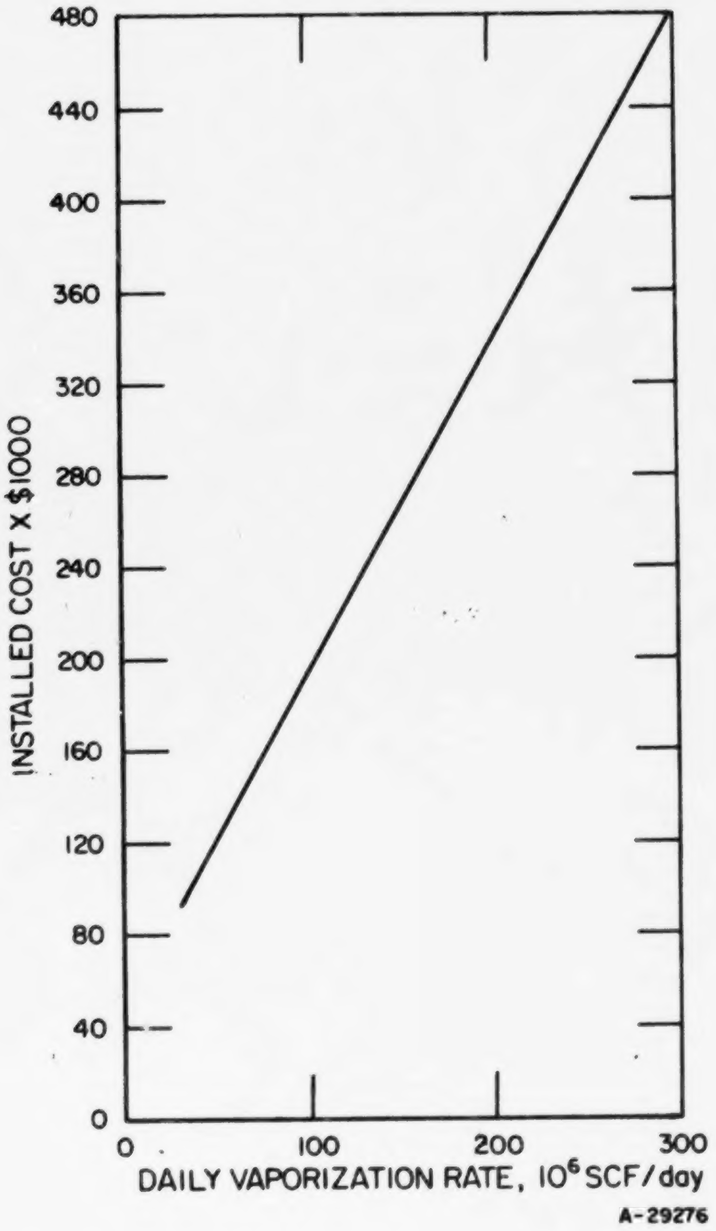


Figure 2. 1. 5. -4. INSTALLED COST OF DIRECT-FIRED AND "SUB-4" TYPE VAPORIZERS

The efficiency of these units is between 70 and 75%. Since most of the loss of fuel heat is in the exhaust gas, efficiency can be improved with an economizer. The heated air is returned to the blower and the efficiency can be increased up to 87%.

2.1.5.1.1.2. Investment Costs

Figure 2.1.5.-4 shows the estimated cost of vaporizers of the size needed for base-load plants. This cost includes the LNG pumps, controls, odorizers, land, and installations costs. Considering the 200 million CF/day plant, which has been our model for the liquefaction system, the cost of the vaporization system is $335,000. It should be noted that sizes above 100 million CF/day gain very little economies of scale. This is to be expected because the primary component is the heat exchanger surface, and this varies directly with vaporization capacity.

Another cost component is the electric motor energy for the combustion air fan. A 250-hp motor is included in the cost. Electrical energy adds significantly to the operating cost.

2.1.5.1.1.3. Operating Costs

In addition to the cost of capital, the other costs are the fuel costs, which can be taken as 40¢/1000 CF of 1100 Btu/CF gas; the power cost can be taken as 5 mills/kWhr. The actual fuel cost is based on an efficiency of 73% in the vaporizer. Electric power is required by the LNG pump (300-psi discharge pressure) and by the combustion air blower.

Referring to Table 2.1.5.-1 it should be noted that the fuel gas is by far the largest cost component. The addition of an economizer would appear to be warranted. Increasing the efficiency to 87% reduces the fuel cost approximately $76,000. The annual cost of the economizer installed is approximately $10,000. This would reduce the vaporization cost to 0.84¢/1000 CF.

Another important point is the effect of LNG pumping on the economics of this system. Liquid pumping is considerably more economical than gas pumping both in capital and energy. For this study, only a moderate pressure was needed because it is assumed that the length of the pipeline between the vaporization plant and the power plant is not more than a few miles.

A large power plant facility would require a minimum of standby capacity because multiple vaporizer units are required. In other words it is conceivable that 15 or 20 individual units are required to meet the total requirements. The cost of providing two standby units is smaller than the accuracy of this estimate; thus this cost has not been included.

2.1.5.1.2. Submerged Combustion Vaporizers

2.1.5.1.2.1. Description

Figure 2.1.5.-5 illustrates the principles of the Sub-X heat exchanger as manufactured by the Thermal Research & Engineering Division of Cosmodyne Corp. The LNG is circulated in stainless steel tubes that are immersed in a water bath. High-pressure fuel and air are burned in the downcomer and the products of combustion flow through the water bubbling up inside the weir. Exhaust gases flow up the stack at almost 100% relative humidity. A highly turbulent motion results from this bubbling that provides high-heat transfer rates. The gas-water mixture is at a lower density than the water outside the weir and as a result water flows under the weir and up around the tubes to further improve heat transfer. The water temperature is 130°F, but the heat flux rates are even higher than the direct-fired heat exchanger because of the wetting action of the water. The manufacturer

Table 2. 1. 5. -1. CALCULATION OF ANNUAL COST OF
LNG VAPORIZATION (Direct-Fired)
Annual Production, 60 billion CF

1. Vaporization Facility Investment	$335,000
2. Capital Cost — Annual	
Principal	$ 22,300
6% Average Interest	$ 10,050
6% Tax Rate and Insurance	$ 20,100
	$ 52,450
3. Fuel Gas	
(15,800 Btu/MCF Vaporized at $0.40/1000 CF)	$472,000
4. Electric Power — Fan, 250 hp	$ 8,700
Pump, 380 hp	$ 13,200
5. Annual Operating Expense, 6% of $335,000	$ 20,100
6. Total Cost	$566,450
7. Unit Cost	0.94¢/1000 CF

states that in the superheat section of the heat exchanger, the heat flux rate is approximately 45,000 Btu/hr/sq ft while in the liquid heating section it is approximately 70,000 Btu/hr/sq ft. This means that the cost of the heat exchanger surface for this method is less than the direct-fired system that has an average heat flux of 20,000 Btu/hr/sq ft.

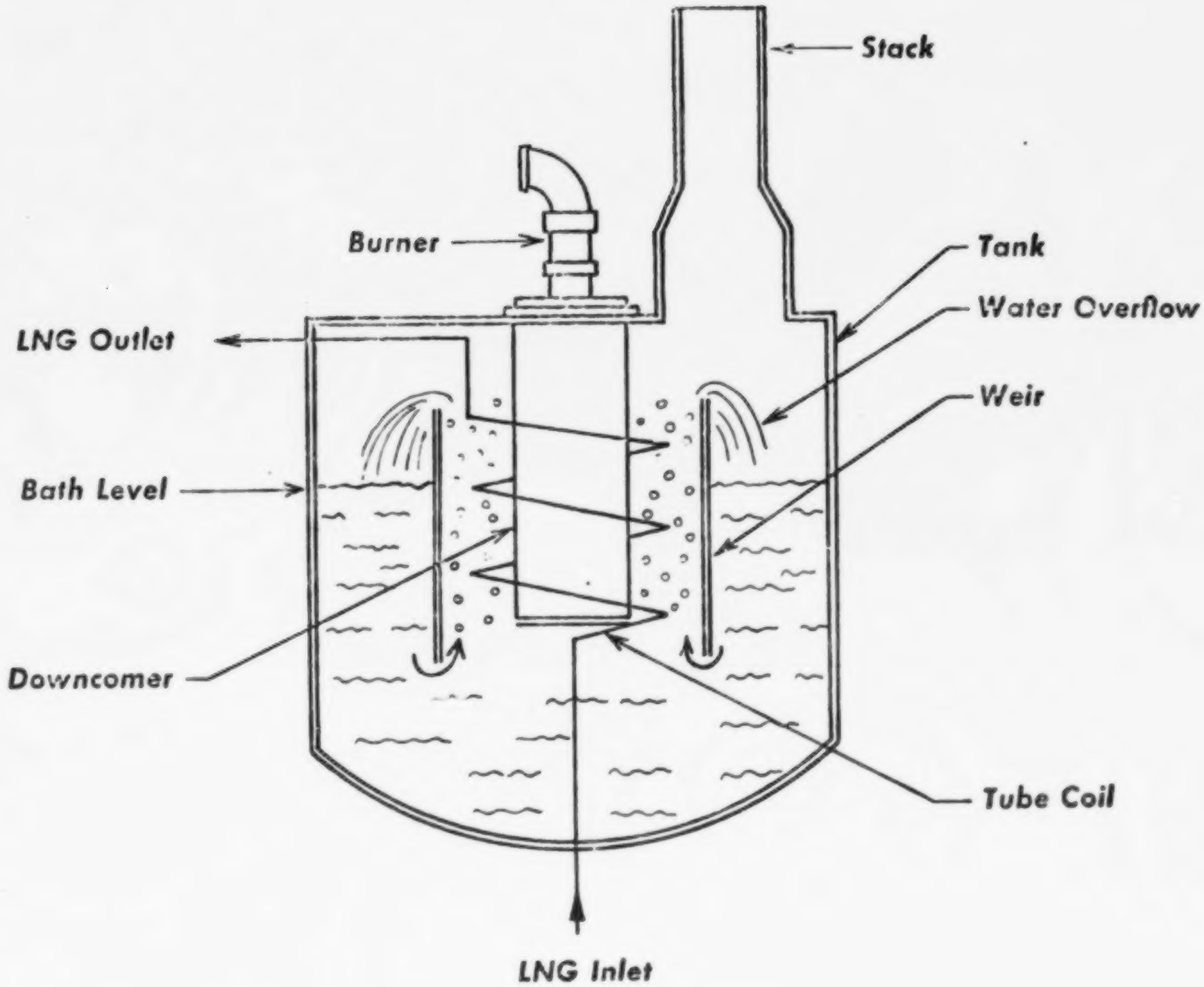

Stack
Burner
Tank
Water Overflow
LNG Outlet
Weir
Bath Level
Downcomer
Tube Coil
LNG Inlet

Figure 2.1.5.-5. SUB-X HEAT EXCHANGER[68]

The efficiency of this system is 90-95% based on the higher heating value of the fuel. Stoichiometric mixtures are used for maximum gas temperature, but the low water temperature produces an excellent low-temperature heat sink and accounts for the high efficiency. The condensation of steam also is a factor improving efficiency. Water produced eliminates the need for "make-up" water. Although LNG leakage is not normal, should a leak occur, the vaporized LNG is in an atmosphere having no oxygen until it reaches the stack. Thus fire hazards are reduced. A similar situation applies to the direct-fired exchanger.

The submerged combustion system has a larger thermal reservoir than the direct-fired exchanger. This tends to provide a cushion for sudden fluctuations in demand. Formation of ice on the tubes is said not to be a problem even at low capacity because of water turbulence.

2. 1. 5. 1. 2. 2. Investment Cost

It has been found that, although the principle of the submerged combustion vaporizers differs greatly from the direct-fired vaporizer, the costs of these two popular systems is quite similar. The investment cost for the 200 million CF/day plant is $335,000. Figure 2. 1. 5.-4 shows the variation of cost with capacity. The higher back pressure on the fan requires that a 200-hp motor be used instead of a 250-hp motor.

2. 1. 5. 1. 2. 3. Operating Costs

The operating cost of this system is very similar to the direct-fired vaporization system with an economizer. There is no need to repeat these figures. The reader is referred to Section 2. 1. 5. 1. 1. 2. for the analysis. The vaporization cost is 0.84¢/1000 CF for this method.

2.1.5.1.3. Indirect-Fired Intermediate Fluid Vaporizers

2.1.5.1.3.1. Description

Figure 2.1.5.-6 illustrates a typical indirect-fired system. The LNG is vaporized and superheated in a heat exchange with a water-glycol mixture. The water-glycol mixture is in turn heated by a natural-gas-fired furnace. The additional cost of the intermediate heat exchange cycle has, according to its proponents, the following advantages.

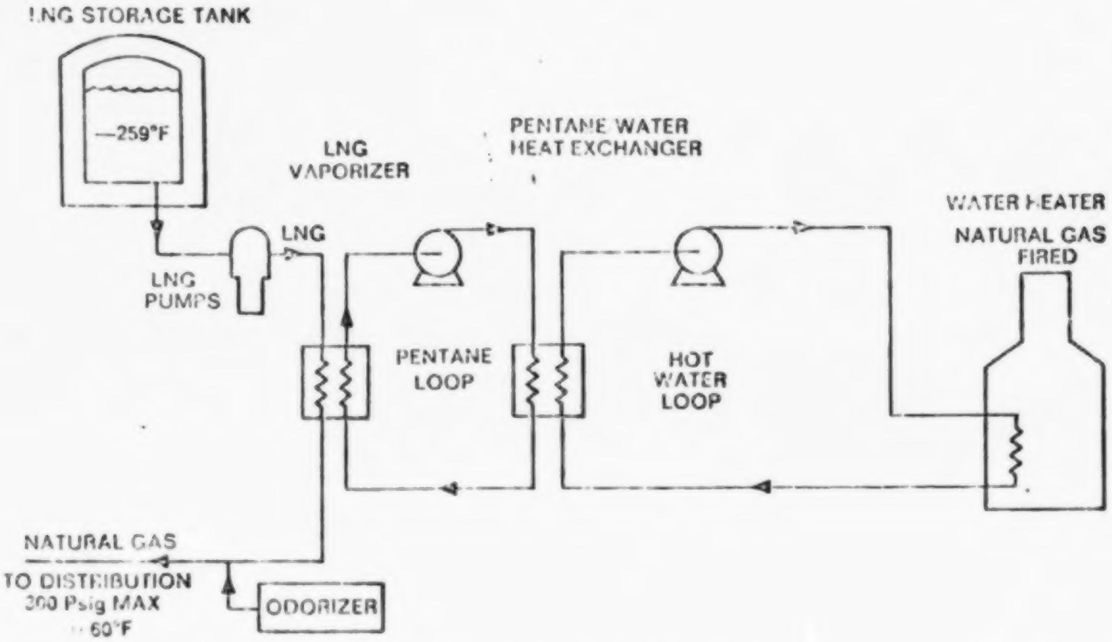

LNG STORAGE TANK
—259°F
LNG VAPORIZER
PENTANE WATER HEAT EXCHANGER
WATER HEATER
NATURAL GAS FIRED
LNG
LNG PUMPS
PENTANE LOOP
HOT WATER LOOP
NATURAL GAS
TO DISTRIBUTION
300 Psig MAX
60°F
ODORIZER

Figure 2.1.5.-6. SIMPLIFIED DIAGRAM OF LNG SENDOUT SYSTEM[29]

1. There is no direct flame contact with LNG piping. This improves the safety of operation.
2. It lends itself well to winterization because the water-glycol system is closed.
3. It has a large turndown capability and can thus operate efficiently over a wide range of heat inputs.

This system has been installed in the first LNG peakshaving plant in the U.S. (Wisconsin Natural Gas) and is also used at the Transco facility. The latter installation is rated at 200 million SCF/day, which, coincidentally, is the size of the plant used in the illustrative examples. The installation at

Canvey Island, England, also uses this method except that propane is the intermediate fluid. At Canvey the two independent systems are each rated at 60 million CF/day.

2.1.5.1.3.2. Investment Costs

It is estimated that the additional cost of heat exchangers, pumps, piping, and controls adds approximately $95,000 to the cost of this system compared to the previously discussed direct-fired and submerged combustion methods. The total cost of this installation including the additional land is estimated at $430,000

2.1.5.1.3.3. Operating Costs

The efficiency of operation will be taken as 73% in consideration of the boiler losses as well as the piping heat loss. The pump power is estimated at approximately 650 hp for the pentane and LNG and glycol-water pumps. Fan horsepower for the boiler is 60 hp. Table 2.1.5.-2 shows a cost analysis for this system.

2.1.5.2. Ambient Air Vaporizers

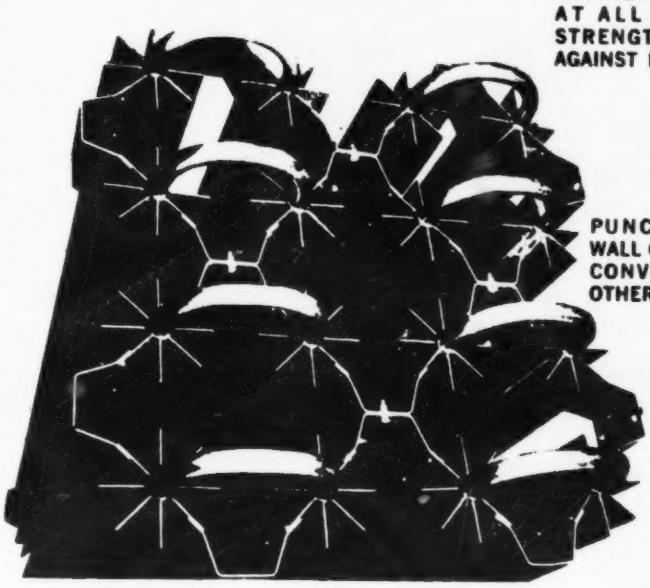
2.1.5.2.1. Description

This type of vaporizer has been used in small applications, but would not be suitable for base-load plants because of the low rate of heat transfer per unit of exchanger area. Figure 2.1.5.-7 shows a typical construction. A very large surface area would be required to vaporize the LNG. An ambient air temperature, winter to summer, would have a significant effect on output, especially in the temperature zone.

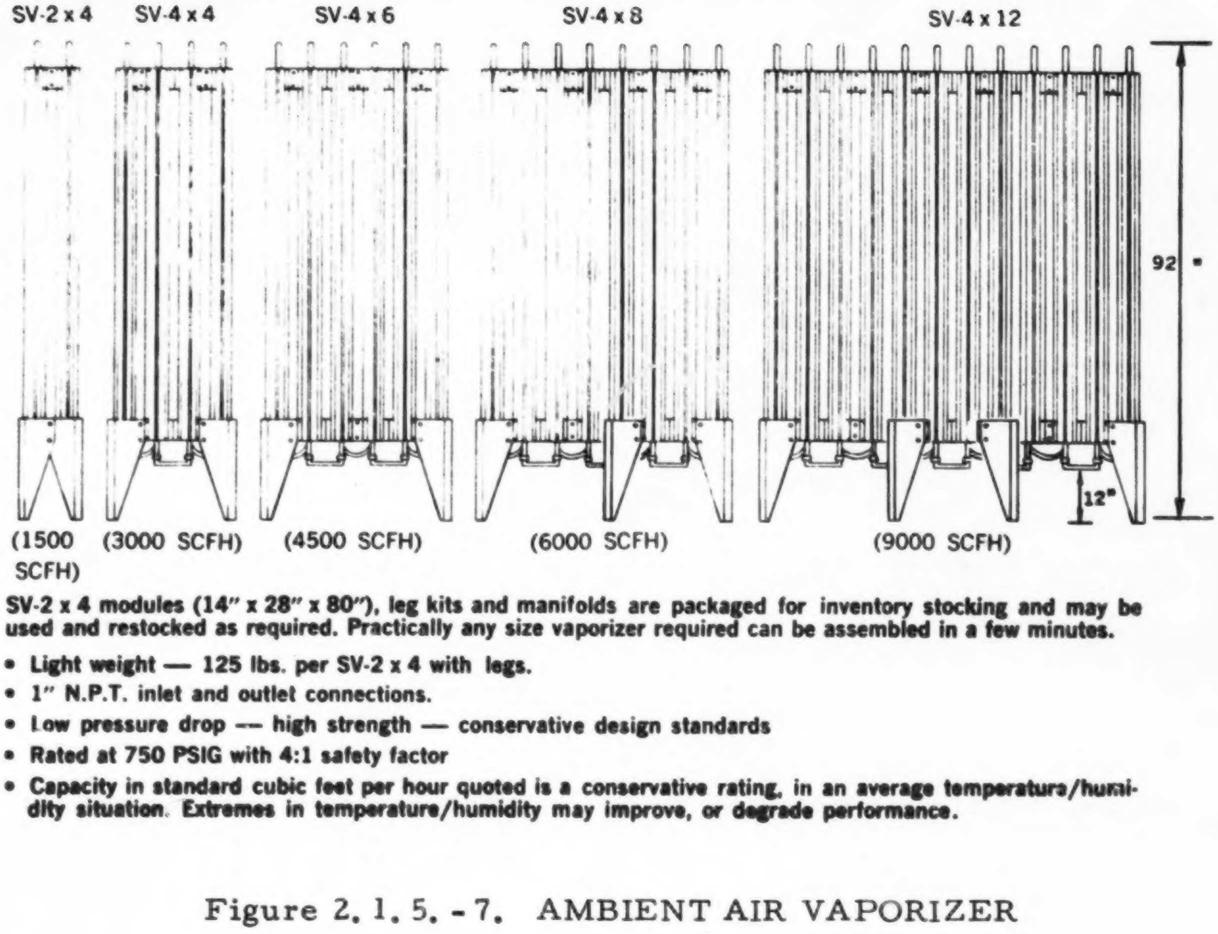
Table 2.1.5.-2. CALCULATION OF UNIT COSTS OF LNG VAPORIZATION
Intermediate Fluid Vaporizers

1. Vaporization Facility Investment	$430,000
2. Capital Cost — Annual	
Principal	$ 28,700
6% Average Interest	$ 12,900
6% Tax Rate and Insurance	$ 25,800
	$ 67,400
3. Fuel Gas	$472,000
4. Electric Power — Pumps, 650 hp	$ 22,600
Fan, 60 hp	$ 2,080
5. Annual Operating Expense, 6% of $430,000	$ 25,800
6. Total Cost	$589,880
7. Unit Cost	0.98¢/1000 CF

COSMODYNE STARFIN VAPORIZER


RIGID MOUNTING CLIP. DOUBLE BOLTED CONSTRUCTION.
COUNTER-BORED CONNECTIONS AT ALL WELDS FOR BEST STRENGTH AND RELIABILITY AGAINST LEAKS.
OPTIMUM INNER/OUTER SURFACE RATIO FOR CRYOGENIC VAPORIZATION
EASY TO ADD OR REMOVE 2 x 4 MODULES TO MEET SPECIFIC CUSTOMER OR APPLICATION REQUIREMENTS.
PUNCHED CLIPS FACILITATE WALL OR TANK MOUNTING. ALSO CONVENIENT FOR SIGNS OR OTHER SYSTEM COMPONENTS.
FULL 2" SPACING BETWEEN STARFINS ALLOWS OPTIMUM CIRCULATION OF AIR AND PREVENTS FROST ARCH-OVER.

HOW TO BUILD A VAPORIZER


SV-2 x 4
SV-4 x 4
SV-4 x 6
SV-4 x 8
SV-4 x 12
92"
12"
(1500 SCFH)
(3000 SCFH)
(4500 SCFH)
(6000 SCFH)
(9000 SCFH)

SV-2 x 4 modules (14" x 28" x 80"), leg kits and manifolds are packaged for inventory stocking and may be used and restocked as required. Practically any size vaporizer required can be assembled in a few minutes.

- **Light weight — 125 lbs. per SV-2 x 4 with legs.**
- **1" N.P.T. inlet and outlet connections.**
- **Low pressure drop — high strength — conservative design standards**
- **Rated at 750 PSIG with 4:1 safety factor**
- **Capacity in standard cubic feet per hour quoted is a conservative rating, in an average temperature/humidity situation. Extremes in temperature/humidity may improve, or degrade performance.**

Figure 2.1.5.-7. AMBIENT AIR VAPORIZER

This type of heat exchanger uses aluminum-plate fins and special cross-section tubing to provide as much area for heat exchange as possible.

Since this design is not applicable to base-load systems, we shall not investigate its investment or operating costs.

2. 1. 5. 3. Unfired Vaporizers (Water)

2. 1. 5. 3. 1. Description

An example of this type of system is shown in Figure 2. 1. 5. -8.

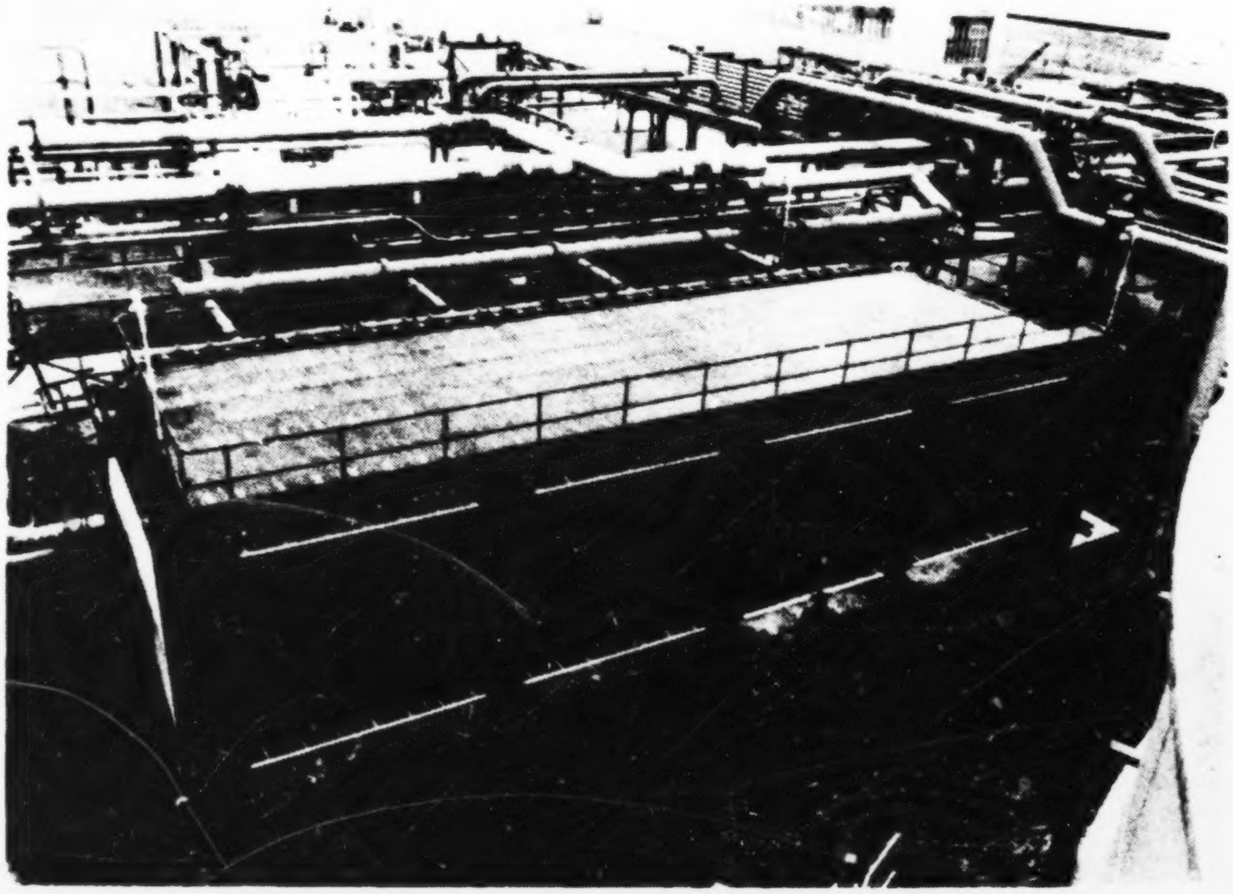


Figure 2. 1. 5. -8. WATER-BATH VAPORIZER USED AT CANVEY ISLAND [44]

This vaporizer uses river water and has a capacity of 96 million CF/day. LNG is pumped into manifolds in the bottom of banks of vertical tubing. The tubing is internally finned and water flows down the outside of the tubes in a thin film. The formation of ice on the tube banks is controlled in thickness by regulating the amount of water flow.

At Canvey Island it is necessary to heat the river water for about 4 weeks of the year, otherwise the water temperature is too low and under heavy demand conditions vaporization capacity is reduced. The vaporized LNG is heated to 32° F and the total water flow required is 1.5 million gph.

2.1.5.3.2. Investment Costs

Water pumping costs are a variable depending upon the degree of salinity of water and the physical arrangements for pumping out the cooled water from the sump. The quantity of water required varies with the season. For purposes of this example we shall assume a 5° F water temperature reduction as a result of heating the LNG and a flow rate of 3 million gph of seawater. It will also be assumed that no supplementary heat will be provided and that the vaporizer will be mounted at sufficient height and close enough to the sea that gravity flow will be sufficient to return the water to the sea. It is estimated that the pump discharge pressure will be 17 psig.

The total cost for this installation including land, controls, pumps, and heat exchangers is approximately $940,000.

2.1.5.3.3. Operating Costs

Two 300-hp electric-motor-driven water pumps will be required to provide full capacity of 200 million CF/day of natural gas. Power cost will be taken at 5 mills/kWhr.

The elimination of fuel cost charges makes a decided improvement in the economics of vaporization by this technique. As can be seen, the cost is less than one-half that of other methods. However, it is still a cost, but, as discussed in later sections, this cold can be a cost-reduction factor rather than a cost-increasing factor, as it is in all the vaporization systems thus far being used.

Table 2. 1. 5. -3 shows an analysis of the operating cost for this type of vaporizer.

Table 2. 1. 5. -3. CALCULATION OF THE ANNUAL COST OF LNG VAPORIZATION WATER VAPORIZER

1. Vaporization Facility Investment	$940, 000
2. Capital Cost – Annual	
Principal	$ 62, 700
6 % Average Interest	$ 28, 200
6 % Tax Rate and Insurance	$ 56, 400
	$147, 300
3. Electric Power – Pumps, 600 hp total	$ 20, 800
4. Annual Operating Expense, 6 % of $940, 000	$ 56, 400
5. Total Cost	$224, 500
6. Unit Cost	0. 37¢/1000 CF

2.2. Technological Trends

2.2.1. Gas Purification System

For bulk purification systems the conventional gas-sweetening processes will continue to find wide acceptance. Molecular sieve systems, utilizing improved materials but probably gaining their principal advantage from the simplification of construction, operation, and regeneration requirements, will find wider acceptance for final gas cleanup. As less pure gases — those which contain a higher percentage of hydrogen sulfide, carbon dioxide, or nitrogen — become the feed source for LNG liquefaction plants, gas cleanup will become an increasingly important part of the cost. It is to be expected that nitrogen separation processes, or at least nitrogen reduction processes, will find wider acceptance as liquefaction plant operators strive for higher heating value LNG.

Liquid absorption processes utilizing alcohol, e.g., the Rectisol Process, will find some degree of acceptance.

2.2.2. Liquefaction Systems

Since liquefaction system cost represents such a large portion of the total cost, there is a continual and extensive research program under way to devise new liquefaction systems. These are primarily designed to reduce the horsepower input requirements and/or heat exchanger surface. A great number of complex schemes have been proposed. The single-refrigerant loop cycle, the so-called autorefrigerated cascade cycle or single-duty compressor cycle, is presently finding increasing acceptance in large plants because of its simplicity of operation. Even though it does have a somewhat lower system efficiency, the reduced complexity of plant construction and operation is a prime advantage in countries where labor skills are low.

2.2.3. Storage System

Double-walled aboveground tanks are presently the most widely accepted method of LNG storage. As demands for greater and greater storage capacity increase, it can be anticipated that storage tank sizes will be increased. The so-called suspended-deck roof construction has simplified tank construction somewhat and, therefore, reduced cost. The search will continue for lower cost materials of construction to eliminate the high cost of cryogenic metal inner tanks. Organic insulation systems, such as urethane or similar materials, will be investigated and should find some degree of acceptance, at least in some specialized designs.

Because of the inherent safety of the belowground storage of LNG, considerable effort is being expended, and will continue to be expended, on developing new concepts in belowground storage vessels. While several designs have been proposed, none have really been fully accepted. Problems of corrosion, water intrusion, buoyancy, and other factors all contribute to making underground metal tank construction somewhat more difficult and costly than aboveground construction. The cryogenic inground system, the so-called "frozen-hole" concept, although temporarily out of favor because of operating problems, may find wide future acceptance because of the inherent lower cost and safety, providing the problems of containment of the LNG can be resolved.

Special in-ground containers, such as atmospheric-pressure and high-pressure caverns, have often been suggested as possible storage alternatives. The one atmospheric pressure cavern (near Tewksbury, Mass.) that has been demonstrated has shown the technical feasibility of building an insulated and lined cavern belowground. Special high-pressure storage caverns constructed at considerable depth below the surface where the hydraulic gradient serves as the sealing medium have also been suggested, but so far technical and economic analyses have not shown justification for such systems.

2.2.4. Transportation Systems

2.2.3.1. Ocean Transportation

The current widespread interest in the membrane tank technology, with its potentially lower cost and better cold utilization factor, will continue to grow. If the tankers now being constructed using the membrane tank technology operate satisfactorily and demonstrate the economy of this system, it is to be expected that a large proportion of the tankers to be built in the future will embody this concept. The principal search will be for lower cost or improved membrane materials and insulation systems. The metals now used are costly and difficult to fabricate. Plastics might offer a lower cost and ultimately better alternative for the containment vessel.

A second interesting possibility is offered by the use of the so-called MLG (modified liquid gas) process for gas transportation. This technique, developed by the Ocean Transport group, uses a ship containing a muliplicity of cargo bottles in which natural gas is transported at a pressure of 200 psig and at a temperature of approximately −175° F. Under these conditions the horsepower requirements for gas compression are said to be about one-half those for the production of LNG. The concept has been tested and demonstrated in a prototype ship and is now ready for commercial application.

High-speed propulsion systems offer a particularly attractive advantage to LNG tankers. Tanker economics are quite different from the conventional oil tanker economics. In the oil tanker trade, using a low-cost oil tanker, it is more economical to use small propulsion systems thereby affording maximum cargo capacity but resulting in slower speeds. On the other hand, an LNG tanker may cost up to 4 times as much for an equivalent carrying capacity, and it may be more economical to develop a high-speed propulsion system, even at the expense of a considerably higher fuel consumption rate. Gas turbines have been suggested as the possible power supply for this type of operation and have speeds up to 30 knots. Speeds of this magnitude would require a special hole design that is quite different from the conventional tanker construction.

Sea barges and ocean-going tugs may have a role to play in LNG transportation. They can enter shallow-draft European ports while today's 300,000-ton and larger oil tankers have only limited access to them. Barges also simplify the structural requirements of design, lend themselves well to mass-production techniques, and have handling flexibility. Powerful pusher tugs could unload barges at a central port, and smaller tugs could distribute the barges along the seacoast as needed. Whether safety requirements would be increased or reduced by barges compared with tankers is a moot point.

Another factor that should be investigated is the possibility that barges can be more economically insulated than tankers.

2.2.4.2. Rail Transport

Technological trends in rail transport will be involved, for the time being as rail transport of LNG begins. Rail tanker designs have been prepared and should be adequate for the near future. At the present time, since tanker design is size-limited, there is little opportunity for increasing the size. Here again, materials of construction could lower the cost and improve the economics of rail transport.

2.2.4.3. Road Transport

The current technological trends are to larger over-the-road tanker trucks. Here the size and carrying capacity will be limited only by the highway regulations, but this does impose a finite upper limit on the transport capacity. Improved insulation systems, such as evacuated powders, or even superinsulation could reduce the space required for insulation and increase the LNG carrying capacity while maintaining the same external tank measurements. Any improvements in cryogenic materials for LNG containment could also improve and reduce tanker costs and improve road transport economics.

2.2.4.4. Barge Design

While several preliminary barge designs have been developed, no extensive analysis has been made to determine the best configuration. Seagoing barges have been proposed as a means of delivering smaller quantities of LNG to areas that could not economically be served by a large deepwater tanker. These barge designs would embody many of the features of the seagoing cryogenic liquid barges now in use, but would be adapted for the special requirements of LNG service. (See Section 2.2.4.1.)

2.2.4.5. Liquid Pipelines

The concept of liquid pipelining has been suggested as a means of reducing natural gas transportation costs. Since it is possible to move approximately 10 times the energy through a pipeline of the same internal diameter if the gas is in the liquid rather than in the gaseous state, there would appear to be some possible economies of pipeline construction. However, these are largely offset by the need for cryogenic materials, insulation, and the outer containment shell construction costs. Also, since the line itself has a constant heat influx, the line must be operated either at supercooled conditions or under two-phase flow. The latter is a highly undesirable condition. Materials suitable for LNG pipeline construction are available. Presently existing insulation systems could be adapted to pipeline service. However, much more work must be done before an economically competitive LNG pipeline system could be developed.

The economics of liquid vs. gas pipelining cannot be examined in a vacuum. The overall system of cold recovery plays an important part in this analysis. Studies made to date indicate that the larger the system the greater the economic advantages of LNG pipelining. Consequently, this system shows promise for a large central power plant such as we have under consideration.

It should be borne in mind that this is not a relatively large cost in the overall power generation system. It is recognized that it is more economical to ship energy by a gas pipeline than by high-voltage power lines. Thus, any saving made by an LNG pipeline vs. a gas line may not be an overriding factor.

2.2.5. Other Pertinent Developments

It may be of value to examine oil, gas, and LNG from the perspective of recent history. Oil first became useful as a national energy resource when its transportation became economical. It was not until high-pressure pipe and field welding became economic realities that oil's place was assured. Exploitation of natural gas in the U.S. was even more dependent on these two mechanical ingredients. Similarly, LNG is awaiting its release from the limitations of present-day technology.

For LNG the great breakthroughs will come in the areas of liquefaction, transportation, and storage. Integrated systems will be developed that will radically change the economics detailed in Section 3.2.4.7. There is little doubt that cold recovery techniques are the key to such a system. These have been mentioned in various places in this report, particularly in Section 2.1.2.5.1.1. under liquid nitrogen exchange.

Just as in the early days of the oil industry technicians were learning to refine oil and to obtain greater markets for different crude oil fractions, so now the peripheral values of LNG importation schemes will need to be developed. The separation of valuable propane and ethane has already been engineered into the LNG receiving port at Barcelona.

IGT has made preliminary investigations regarding the combination of nuclear power plants with oxygen and natural gas magnetohydrodynamic-(MHD)-fueled topping units. These would operate with an LNG importation scheme. Such a hybrid plant appears to overcome many of the thermal pollution problems of nuclear energy and makes significant use of the cold from LNG to produce the oxygen required by the MHD portion. The liquid nitrogen by-product is used in the backhaul and provides refrigeration at the wellhead to liquefy the natural gas. This concept is a supplement to the work recently reported by Brookhaven National Laboratory (BNL 12319-R and BNL 12569). This sort of plan is also compatible with desalination schemes.

Although multiple-function importation schemes combined with hybrid fossil-nuclear fueled MHD power cycles may at first appear rather complex, we must suppose that multiple fractionation of crude oil and balancing market requirements must have appeared very complex to the oil industry's early competitors. Even though the process is indeed complex, it is also highly profitable and even necessary.

Finally, in terms of world resources of petroleum and natural gases, it is interesting to note that the ratio of proved oil and gas reserves is approximately 2:1; that is, the heat content of world oil reserves (proved only) is twice as great as for gas.[24] When we consider that the search for oil predates the search for gas and that, in many areas of the world, gas has no commercial value (it is flared to the atmosphere), the abundance of natural gas proved reserves is surprisingly high. With improvement in LNG economics and technology, it can be expected that natural gas, having first passed through the LNG stage, will make an increasing contribution to world and U. S. energy supplies.

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Section 3.0. Comparison of Power Generating Costs Using Different Fossil Fuels

Clean-burning natural gas and LNG-fired electric generation costs are compared to the cost of using coal and oil when the latter require flue gas desulfurization. The final comparison is made on the electrical bus bar cost, including all capital charges, fuel costs, and other operating costs for each alternative fuel.

Imported LNG costs have been estimated for both the East and West Coasts from several likely sources and at several foreign gas purchase costs. Alternative sources that could be competitive with high-sulfur fuels for electric power generation are identified for both coasts.

3.1 Metropolitan Area Characteristics

To help evaluate the potential for coastal LNG power generation blocks of the magnitudes contemplated, data were collected and organized on a state-by-state basis (with some deviations), which covered installed steam electric capacity, net annual generation, and fuels used during recent years. The two coastal groupings of these data for 1960 through 1966 are presented in Tables 3.1.-1 through 3.1.-7.

For the same geographic subdivisions, recent plans for annual steam electric capacity additions through 1972 are itemized in Table 3.1.-8. These planned additions are also broken into fossil-fueled and nuclear-fueled steam electric plant categories.

By combining certain data in Tables 3.1.-7 and 3.1.-8, we show in Table 3.1.-9 the cumulative installed steam electric capacity in the two coastal areas for 1967 through 1972. (These figures are unadjusted for plant retirements or possible future additions, delays, or cancellations.)

To assist in visualizing the magnitude of a 10,000-MW block of steam electric generating capacity, it is noted from Table 3.1.-7 that the total capacity installed in New York City in 1966 was 8909.3 MW. California had 15,848.7 MW of steam electric capacity. Table 3.1.-10, again based on 1966 levels, details a hypothetical grouping of existing plants to compose a block of approximately 10,000 MW capacity. It includes most of the capacity of Consolidated Edison Company of New York, Inc. (excluding the Indian Point nuclear plant and several combination district heating and power production plants), essentially all of the capacity of Long Island Lighting Company, and two of the larger stations owned by Public Service Electric and Gas Company. Based on historic trends, the demands of the region serviced by this capacity might be expected to approximately double by 1976 and quadruple by 1986.

The growth of steam-electric generation capacity along the East and West Coasts since 1960 has been projected to 1985 (Figure 3. 1. -1.). Tables 3. 1. -1 through 3. 1. -9 were the latest available source of data for the 1960-1966 period. The projections were scaled to energy growth and capacity factor forecasts by Ebasco Services, Inc.,[6] that were based on U. S. Census divisions of similar, though not identical, geographic extent.

In 1975, the entire East Coast area, as selected, will have approximately 110,000 MW of steam-electric generating capacity, including both nuclear- and fossil-fueled units but excluding hydroelectric capacity. Thus, there will be 11 times as much installed capacity along the East Coast as in the 10,000-MW metropolitan area generating block arbitrarily chosen for examination in this study. Projections indicate that 1985 East Coast capacity will reach 215,000 MW, or the equivalent of more than 21 such blocks.

In the selected West Coast area, there will be about 44,400 MW (4-1/2 blocks) in 1975, and 100,000 MW(10 blocks) in 1985.

If, by definition, a 10,000-MW "block" of power is restricted to the demand of a given greater metropolitan area, then the number of such blocks on each coast is greatly diminished. In 1966, for instance, only the greater New York City area would have qualified. By comparison, the entire state of Massachusetts had only 3800 MW of steam-electric capacity. Pennsylvania had slightly over 10,000 MW, but Philadelphia had only 3200 MW. California had a total of 15,848 MW. The Los Angeles-San Bernardino metropolis did not have a total of 10,000 MW, although it closely approached that amount.

Looking to the future, however, the opportunity for finding 10,000-MW blocks in coastal, metropolitan areas will be increasing rapidly. Figure 3.1.-2 indicates how rapidly a 10,000-MW block on each coast in 1966 will be growing. This assumes that the regional growth rates indicated in Figure 3.1.-1 apply to individual cities. For example, the metropolitan New York City block will more than double by 1975, and again double by 1985. The expected West Coast growth rate is even higher. If we encompass enough area around Los Angeles to have a 10,000-MW demand in 1966, it will have 27,500 MW by 1975 and 61,900 MW by 1985. Conversely, if we visualize a 10,000-MW block in 1975, it was only 3630 MW in size in 1966, but will be 22,500 MW by 1985.

The projections cited to this point indicate that the metropolitan areas immediately surrounding New York City, Los Angeles, and possibly San Francisco might be the only ones with as much as 10,000 MW of steam-electric capacity in 1975. If this is to be the case, however, one should keep in mind the qualifications on which such projections would be based: The

Table 3.1.-1. STEAM ELECTRIC PLANTS – 1960 CAPACITY, GENERATION, AND FUEL USAGE

Selected East Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used: Coal, 10^9 Btu	Coal, %	Oil, 10^9 Btu	Oil, %	Gas, 10^9 Btu	Gas, %	Total, 10^9 Btu
Maine	274.0	966.8	486	4	11,019	96	--	--	11,505
N.H.	361.8	1,075.9	2,494	21	9,401	79	--	--	11,895
Mass.	2,731.1	12,322.6	64,377	46	65,014	47	10,340	7	139,731
R.I.	404.8	1,648.4	16,131	79	3,940	19	366	2	20,437
Conn.	1,707.0	7,379.1	73,934	86	10,061	12	1,660	2	85,655
N.Y.C.	5,957.6	24,382.0	174,616	59	63,802	22	55,497	19	293,915
N.J.	4,545.7	17,320.3	91,376	50	63,958	35	26,932	15	182,266
D.C.	530.0	972.8	12,247	99	120	1	--	--	12,367
Phila.	2,920.8	12,196.0	104,148	81	19,875	15	4,959	4	128,982
Del.	407.0	2,209.6	17,512	84	130	--	3,286	16	20,928
Md.	1,734.8	7,729.4	81,792	99	453	1	--	--	82,245
Va.	3,149.6	17,002.0	160,835	99	297	--	1,388	1	162,520
N.C.	3,425.3	14,991.4	147,902	100	92	--	122	--	148,116
S.C.	1,243.0	6,446.2	40,583	62	111	--	24,492	38	65,186
Ga.	1,575.0	8,389.1	65,133	71	247	--	26,255	29	91,635
Fla.	3,917.3	16,572.5	27,190	14	78,318	42	82,707	44	188,215
Total	34,884.8	151,604.1	1,080,756	66	326,838	20	238,004	14	1,645,598

Selected West Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Coal, 10^9 Btu	Coal, %	Oil, 10^9 Btu	Oil, %	Gas, 10^9 Btu	Gas, %	Total, 10^9 Btu
Wash.	186.8	(–) 2.0	--	--	96	100	--	--	96
Oreg.	126.0	8.0	--	--	9	1	740	99	749
Calif.	9,063.2	45,751.9	--	--	161,053	32	337,017	68	498,070
Total	9,376.0	45,757.9	--	--	161,158	32	337,757	68	498,915

Data Source: National Coal Association, Steam-Electric Plant Factors – 1960. Washington, D.C., 1961.

Table 3.1.-2. STEAM ELECTRIC PLANTS – 1961
CAPACITY, GENERATION, AND FUEL USAGE

Selected East Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used						
			Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Maine	274.0	1,208.4	--	--	14,306	100	--	--	14,306
N.H.	393.4	1,546.2	9,335	58	6,882	42	--	--	16,217
Mass.	2,986.4	12,918.9	73,946	50	65,526	44	8,183	6	147,655
R.I.	404.8	1,511.3	14,563	78	3,826	20	409	22	18,798
Conn.	1,842.0	7,960.8	82,013	91	8,559	9	378	--	90,950
N.Y.C.	6,625.5	25,889.5	168,919	55	82,438	26	58,019	19	309,376
N.J.	4,907.1	18,952.1	113,586	59	53,912	28	24,034	13	191,532
D.C.	533.8	820.7	10,732	99	87	1	--	--	10,819
Phila	2,920.8	13,140.4	113,982	84	19,866	15	1,663	1	135,511
Del.	495.9	2,271.1	17,616	84	106	1	3,241	15	20,963
Md.	1,984.0	8,428.7	86,754	99	498	1	65	--	87,317
Va.	3,679.2	17,987.6	168,745	99	734	--	1,593	1	171,072
N.C.	3,692.0	16,372.2	159,483	99	53	--	1,007	1	160,543
S.C.	1,243.0	7,426.9	50,886	69	142	--	23,012	31	74,040
Ga.	1,894.7	7,842.9	69,227	80	458	--	17,303	20	86,988
Fla.	4,751.1	19,026.2	36,227	17	90,565	43	84,078	40	210,870
Total	38,627.7	163,303.9	1,176,014	67	347,958	20	222,985	13	1,746,957
Selected West Coast Service Area									
Wash.	136.8	(–) 2.8	--	--	59	100	--	--	59
Oreg.	111.0	(–) 2.6	--	--	7	1	613	99	620
Calif.	9,958.3	53,972.7	--	--	153,991	27	423,847	73	577,838
Total	10,206.1	53,967.3			154,057	27	424,460	73	578,517

Data Source: National Coal Association, Steam-Electric Plant Factors – 1961. Washington, D.C., 1962.

Table 3.1.-3. STEAM ELECTRIC PLANTS – 1962
CAPACITY, GENERATION, AND FUEL USAGE

Selected East Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used						
			Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Maine	276.3	1,396.1	--	--	16,566.8	100	--	--	16,566.8
N.H.	363.4	1,540.6	9,674.1	60	6,568.1	40	--	--	16,242.2
Mass.	3,019.5	13,451.6	76,972.7	52	67,098.7	43	8,698.1	5	152,769.5
R.I.	404.8	1,478.9	14,472.8	80	3,693.1	19	362.8	1	18,528.7
Conn.	1,834.4	8,779.2	83,308.3	91	9,000.0	9	386.9	--	97,695.2
N.Y.C.	7,226.3	26,905.6	149,903.5	48	98,828.6	31	67,811.7	21	316,543.8
N.J.	5,445.2	19,933.0	115,602.1	59	54,519.6	27	28,873.8	14	198,995.5
D.C.	533.8	705.6	9,346.0	99	107.0	1	--	--	9,453.0
Phila.	2,978.0	13,614.0	118,896.4	87	18,279.3	13	963.5	--	138,139.2
Del.	547.4	1,933.1	16,216.7	82	307.5	1	3,464.5	17	19,988.7
Md.	2,180.5	9,619.4	97,662.9	100	481.5	--	69.7	--	98,214.1
Va.	4,128.3	20,266.4	190,171.3	100	914.7	--	1,596.1	--	192,682.1
N.C.	4,053.5	18,131.5	173,731.9	99	145.2	--	2,784.5	1	176,661.6
S.C.	1,456.8	8,016.8	56,513.2	72	287.8	--	22,506.2	28	79,307.2
Ga.	1,933.1	7,502.4	65,637.3	80	201.7	--	16,931.4	20	82,770.4
Fla.	5,292.8	21,798.5	39,646.8	17	110,060.7	46	87,968.0	37	237,675.5
Total	41,674.1	175,072.7	1,222,756.0	66	387,060.3	21	242,417.2	13	1,852,233.5

Selected West Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Coal 10^9 Btu	Coal %	Oil 10^9 Btu	Oil %	Gas 10^9 Btu	Gas %	Total 10^9 Btu
Wash.	188.5	(–) 2.0	--	--	78.4	100	--	--	78.4
Oreg.	111.0	17.9	--	--	150.5	22	538.7	78	689.2
Calif.	10,823.4	50,491.7	--	--	114,123.1	22	426,450.1	78	540,573.2
Total	11,122.9	50,507.6	--	--	114,352.0	21	426,988.8	79	541,340.8

Data Source: National Coal Association, Steam-Electric Plant Factors – 1962, Washington, D.C., 1963.

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Table 3.1.-4. STEAM ELECTRIC PLANTS – 1963
CAPACITY, GENERATION, AND FUEL USAGE

Selected East Coast Service Area

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used						
			Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Maine	288.8	1,271.2	--	--	15,097.6	100	--	--	15,097.6
N.H.	367.2	1,526.6	9,635.2	60	6,470.2	40	--	--	16,105.4
Mass.	3,144.5	14,183.5	83,059.9	52	64,201.9	41	11,045.7	7	158.307.5
R.I.	397.6	1,396.0	12,860.0	73	4,481.7	25	317.9	2	17,659.6
Conn.	2,094.5	9,444.2	92,370.3	88	12,705.4	12	270.7	--	105,346.4
N.Y.C.	7,978.8	28,263.3	139,849.6	43	103,180.9	32	79,749.1	25	322,779.6
N.J.	5,386.3	21,780.9	125,282.2	58	60,404.9	28	30,102.8	14	215,789.9
D.C.	533.8	704.5	9,371.2	99	100.8	1	--	--	9,472.0
Phila.	3,205.2	13,511.5	117,072,8	83	20,599.6	15	3,148.7	2	140,821.1
Del.	547.4	2,131.9	18,391.4	83	362.8	2	3,405.7	15	22,159.9
Md.	2,389.9	11,340.9	113,006.9	100	483.7	--	90.6	--	113,581.2
Va.	4,115.8	20,129.1	195,218.2	99	838.3	--	1,733.6	1	197,790.1
N.C.	4,053.5	20,541.7	196,998.1	98	142.9	--	2,968.4	2	200,109.4
S.C.	1,456.8	8,880.0	64,183.4	73	311.0	--	23,208.2	27	87,702.6
Ga.	2,232.3	9,087.8	93,692.6	97	247.5	--	2,990.9	3	96,931.0
Fla.	5,863.6	24,065.6	43,271.6	17	121,554.7	47	94,160.0	36	258,986.3
Total	44,056.0	188,258.7	1,314,263.4	66	411,183.9	21	253,192.3	13	1,978,639.6
Selected West Coast Service Area									
Wash.	188.5	(–) 3.0	--	--	36.9	100	--	--	36.9
Oreg.	111.0	20.8	--	--	51.1	8	621.8	92	672.9
Calif.	12,329.7	52,537.5	--	--	101,907.3	19	425,599.2	81	527,506.5
	12,629.2	52,555.3			101,995.3	19	426,221.0	81	528,216.3

Data Source: National Coal Association, Steam-Electric Plant Factors – 1963. Washington, D.C., 1964.

Table 3.1.-5. STEAM ELECTRIC PLANTS – 1964
CAPACITY, GENERATION, AND FUEL USAGE

			Fuel Used						
	Installed Capacity, MW	Net Generation, 10^6 kWhr	Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Selected East Coast Service Area									
Maine	316.0	1,654.4	--	--	19,630.6	100	-	--	19,630.6
N.H.	367.2	1,623.6	10,458.4	61	6,579.7	39	--	--	17,038.1
Mass.	3,358.7	16,082.9	96,239.0	55	67,071.9	38	12,457.4	7	175,768.3
R.I.	397.6	1,174.2	10,557.9	68	4,549.4	29	465.6	3	15,572.9
Conn.	2,333.9	10,235.8	97,868.5	12	13,378.2	88	390.0	--	111,636.7
N.Y.C.	7,938.8	30,805.6	166,245.0	47	108,618.6	31	76,850.5	22	351,714.1
N.J.	6,004.3	24,261.3	143,079.8	59	69,325.3	29	29,210.0	12	241,615.1
D.C.	533.8	865.7	11,513.2	99	138.6	1	--	--	11,651.8
Phila.	3,205.2	14,262.7	125,147.0	82	24,192.5	16	3,610.4	2	152,949.9
Del.	547.4	2,435.8	21,444.4	84	397.7	2	3,660.1	14	25,502.2
Md.	2,761.2	12,677.2	126,583.4	99	818.2	1	64.7	--	127,466.3
Va.	4,474.8	22,134.4	208,320.7	99	1,058.8	--	1,927.9	1	211,307.4
N.C.	4,260.1	22,191.1	214,045.5	99	171.8	--	2,522.4	1	216,739.7
S.C.	1,592.8	8,340.7	62,433.7	76	195.3	--	19,477.2	24	82,106.2
Ga.	2,694.7	10,111.9	101,176.1	97	328.6	--	2,701.5	3	104,206.2
Fla.	6,757.5	26,165.2	46,694.9	17	142,850.8	51	90,690.7	32	280,236.4
Total	47,544.0	205,022.5	1,441,807.5	67	459,306.0	21	244,028.4	11	2,145,141.9
Selected West Coast Service Area									
Wash.	188.5	–(2.9)	--	--	18.3	100	--	--	18.3
Oreg.	111.0	4.7	--	--	12.9	4	342.1	96	355.0
Calif.	13,756.8	63,797.2	--	--	108,003.1	17	520,595.9	83	628,599.0
Total	14,056.3	63,799.0			108,034.3		520,938.0		628,972.3

Data Source: National Coal Association, Steam-Electric Plant Factors – 1964. Washington, D.C., 1965.

Table 3.1.-6. STEAM ELECTRIC PLANTS – 1965 CAPACITY, GENERATION, AND FUEL USAGE

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used						
			Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Selected East Coast Service Area									
Maine	449.1	2,335.9	--	--	27,014.7	100	--	--	27,014.7
N.H.	367.2	1,720.0	9,626.8	52	9,026.9	48	--	--	18,653.7
Mass.	3,725.1	18,281.8	105,701.3	54	78,516.2	40	12,286.6	6	196,504.1
R.I.	398.1	1,264.5	11,041.9	67	4,969.7	30	509.4	3	16,521.0
Conn.	2,322.7	11,391.3	106,273.5	87	16,171.6	13	333.2	--	122,778.3
N.Y.C.	8,944.3	31,749.9	157,738.7	43	134,619.7	37	75,189.9	20	367,548.3
N.J.	6,013.9	26,789.9	173,065.7	65	71,404.8	27	23,037.5	8	267,508.0
D.C.	533.8	560.7	7,895.9	98	126.0	2	--	--	8,021.9
Phila.	3,205.2	15,132.7	132,879.8	85	24,282.0	15	37.3	--	157,298.1
Del.	548.0	2,570.0	21,957.5	82	210.6	1	4,592.0	17	26,760.1
Md.	3,125.0	15,936.9	158,071.9	99	991.3	1	56.4	--	159,119.6
Va.	4,474.3	21,783.4	212,880.2	99	377.2	--	1,747.2	1	215,004.6
N.C.	4,610.1	25,745.0	245,198.3	99	139.8	--	3,078.1	1	248,416.2
S.C.	1,592.8	8,930.6	67,116.0	77	323.3	--	19,964.2	23	87,133.5
Ga.	3,097.4	12,958.8	130,867.8	99	317.2	--	819.8	1	132.004.8
Fla.	7,985.3	29,509.9	55,173.8	18	172,070.8	55	84,236.7	27	311,481.3
Total	51,392.8	226,661.3	1,595,588.1	68	540,561.8	23	225,618.3	10	2,361,768.2
Selected West Coast Service Area									
Wash.	188.5	--	--	--	23.0	100	--	--	23.0
Oreg.	111.0	--	--	--	6.4	5	121.4	95	127.8
Calif.	14,948.7	61,572.0	--	--	111,644.8	19	489,672.8	81	601,137.6
Total	15,248.2	61,572.0			111,674.2		489,794.2		601,468.4

Data Source: National Coal Association, Steam-Electric Plant Factors – 1965. Washington, D.C., 1966.

Table 3.1.-7. STEAM ELECTRIC PLANTS – 1966
CAPACITY, GENERATION, AND FUEL USAGE

	Installed Capacity, MW	Net Generation, 10^6 kWhr	Fuel Used						
			Coal		Oil		Gas		Total
			10^9 Btu	%	10^9 Btu	%	10^9 Btu	%	10^9 Btu
Selected East Coast Service Area									
Maine	436.6	2,400.8	--	--	27,106.6	100	--	--	27,106.6
N.H.	367.2	1,726.3	8,456.5	45	10,211.0	55	--	--	18,667.5
Mass.	3,793.1	19,717.3	95,262.8	45	106,013.5	50	9,318.2	5	210,594.5
R.I.	365.1	1,284.0	10,243.3	62	6,129.8	37	232.8	1	16,605.9
Conn.	2,235.5	12,363.1	113,885.6	84	21,902.6	16	363.3	--	136,151.5
N.Y.C.	8,909.3	33,776.1	136,927.5	35	178,657.0	46	71,894.7	19	387.479.2
N.J.	5,991.4	28,749.8	180,255.9	61	93,809.1	32	22,060.2	7	296,125.2
D.C.	533.8	916.5	13,001.3	98	201.6	2	--	--	13,202.9
Phila.	3,205.2	16,006.5	128,644.4	76	41,655.8	24	--	--	170,300.2
Del.	725.3	3,382.6	30,505.9	87	320.5	1	4,083.6	12	34,910.0
Md.	3,484.0	17,321.3	174,710.1	99	1,100.6	1	73.8	--	175,884.5
Va.	4,441.3	22,543.5	216,548.1	99	420.5	--	1,428.0	1	218,396.6
N.C.	5,371.0	30,759.4	294,061.6	99	196.3	--	2,788.7	1	297,046.6
S.C.	1,665.3	9,665.5	75,275.3	79	756.1	1	19,091.3	20	95,122.7
Ga.	3,097.4	14,769.1	149,905.9	100	190.8	--	410.8	--	150,507.5
Fla.	8,748.3	33,631.8	70,138.0	--	188,629.6	--	95,786.6	--	354,554.2
Total	53,369.8	249,013.6	1,697,822.2	65.2	677,301.4	26	227,532.0	8.8	2,602,655.6
Selected West Coast Service Area									
Wash.	197.5	83.1	--	--	1,356.5	100	--	--	1,356.5
Oreg.	111.0	13.5	--	--	367.5	74	130.4	26	497.9
Calif.	15,848.7	74,764.9	--	--	127,665.7	18	595,829.9	82	723,495.6
Total	16,157.2	74,861.5	--	--	129,389.7	18	595,960.3	82	725,350.0

Data Source: National Coal Association, Steam-Electric Plant Factors – 1966. Washington, D.C., 1967.

Table 3.1.-8. SCHEDULED ADDITIONS OF STEAM ELECTRIC GENERATION CAPACITY

	East Coast												
	Fossil						Nuclear						Total Steam-Electric
	1967	1968	1969	1970	1971	1972	1967	1968	1969	1970	1971	1972	
Maine												830	830
N.H.		333											333
Mass.	400		630								625		1,655
R.I.													--
Conn.		400			400	400	570		652.1				2,422.1
N.Y.C.	35	515							873		965		2,388
N.J.			25					515				1050	1,590
D.C.		275											275
Phila.							40				1065		1,105
Del.				167									167
Md.				556	558	1275							2,389
Va.			669								780	780	2,229
N.C.		650	671	671							841	841	3,674
S.C.	215		320	375						663			1,573
Ga.		490	500	505	838								2,333
Fla.	1450.5	411	921	757	934	730				688.5	688.5	855	7,435.5
Total	2100.5	3074	3736	3031	2730	2405	610	515	1525.1	1351.5	4964.5	4356	30,398.6
	West Coast												
Wash.					700								700
Oreg.													
Calif.	2064.7	837			808	2405	375					1060	7,549.7
Total	2064.7	837			1508	2405	375					1060	8,249.7
													38,648.3

Data Sources: 1967 Data – Derived from "41st Semi-Annual Electric Power Survey," Edison Electric Institute Publication No. 67-30, April 1967

1968 Data – Derived from a special, unpublished compilation of scheduled additions prepared by Electrical World, covering plans through July 8, 1968.

Table 3.1.-9. INSTALLED GENERATION CAPACITY IN SELECTED COASTAL SERVICE AREAS

	East Coast	West Coast	Both Coasts
Installed Capacity, Dec. 31, 1966, MW	53,369.8	16,157.2	69,527.0
1967 Additions – Fossil	2,100.5	2,064.7	4,165.2
1967 Additions – Nuclear	610.0	375.0	985.0
Installed Capacity 1967	56,080.3	18,596.9	74,677.2
1968 Additions – Fossil	3,074.0	837.0	3,911.0
1968 Additions – Nuclear	515.0	--	515.0
Installed Capacity 1968	59,669.3	19,433.9	79,103.2
1969 Additions – Fossil	3,736.0	--	3,736.0
1969 Additions – Nuclear	1,525.1	--	1,525.1
Installed Capacity 1969	64,930.4	19,433.9	84,364.3
1970 Additions – Fossil	3,031.0	--	3,031.0
1970 Additions – Nucl	1,351.5	--	1,351.5
Installed Capacity 1970	69,312.9	19,433.9	88,746.8
1971 Additions – Fossil	2,730.0	1,508.0	4,238.0
1971 Additions – Nuclear	4,964.5	--	4,964.5
Installed Capacity 1971	77,007.4	20,941.9	97,949.3
1972 Additions – Fossil	2,405.0	2,405.0	4,810.0
1972 Additions – Nuclear	4,356.0	1,060.0	5,416.0
Installed Capacity 1972	83,768.4	24,406.9	108,175.3

Data Sources: National Coal Association, Steam-Electric Plant Factors – 1966. Washington, D.C., 1967.
1967 Data – Derived from "41st Semi-Annual Electric Power Survey," Edison Electric Institute Publication No. 67-30, April 1967.
1968 Data – Derived from a special, unpublished compilation of scheduled additions prepared by Electrical World, covering plans through July 8, 1968.

Table 3.1.-10. COMPOSITION OF A HYPOTHETICAL BLOCK OF 10,000 MW CAPACITY OF EAST COAST – 1966

Con Edison	Plant	MW	Net Gen, 10^6 kWhr	Fuel*	Btu Consumption, %		
					Coal	Oil	Gas
N.Y.C.	Astoria	1,550.6	6,599.9	C(P) OG	85	--	15
N.Y.C.	East River	833.7	1,944.9	C(P)O	50	46	4
N.Y.C.	59th St.	149.5	499.0	C(P)O	61	39	--
N.Y.C.	Hell Gate	611.3	1,580.3	C(P)OG	--	79	21
N.Y.C.	Hudson Ave.	845.0	2,213.8	C(P)O	16	84	--
N.Y.C.	Kent Ave.	107.5	168.0	C(P)OG	74	--	26
N.Y.C.	Arthur Kill	376.2	2,246.2	C(P)O	94	6	--
N.Y.C.	Ravenswood	1,827.7	8,195.8	COG	--	72	28
N.Y.C.	74th St.	269.0	570.1	C(P)O	92	8	--
N.Y.C.	Sherman Creek	216.5	538.1	C(P)O	75	16	9
N.Y.C.	Waterside	712.3	2,184.0	C(P)OG	32	2	66
		7,499.3	26,740.1				
L.I. Lighting Co.							
Island Park, N.Y.	E. F. Barrett	375.0	2,342.3	C(P)OG	33	55	12
Far Rockaway, N.Y.	Far Rockaway	148.2	534.0	C(P)OG	--	96	4
Glenwood Landing, N.Y.	Glenwood	400.3	1,441.4	C(P)OG	--	54	46
Northport, N.Y.	Northport	19.5	1.4	O	--	100	--
Port Jefferson, N.Y.	Port Jefferson	467.0	2,716.9	C(P)O	16	84	--
		1,410.0	7,036.0				
Public Service Electric & Gas Co.							
Jersey City, N.J.	Hudson	454.8	2,543.8	CO	99	1	--
Ridgefield, N.J.	Bergen	650.4	3,903.5	C(P)G	75	--	25
		1,105.2	6,447.3				
Total		10,014.5	40,223.4				

* C = Coal; (P) = Pulverized; O = Oil; G = Gas

Data Source: National Coal Association, Steam-Electric Plant Factors – 1966. Washington, D.C., 1967.

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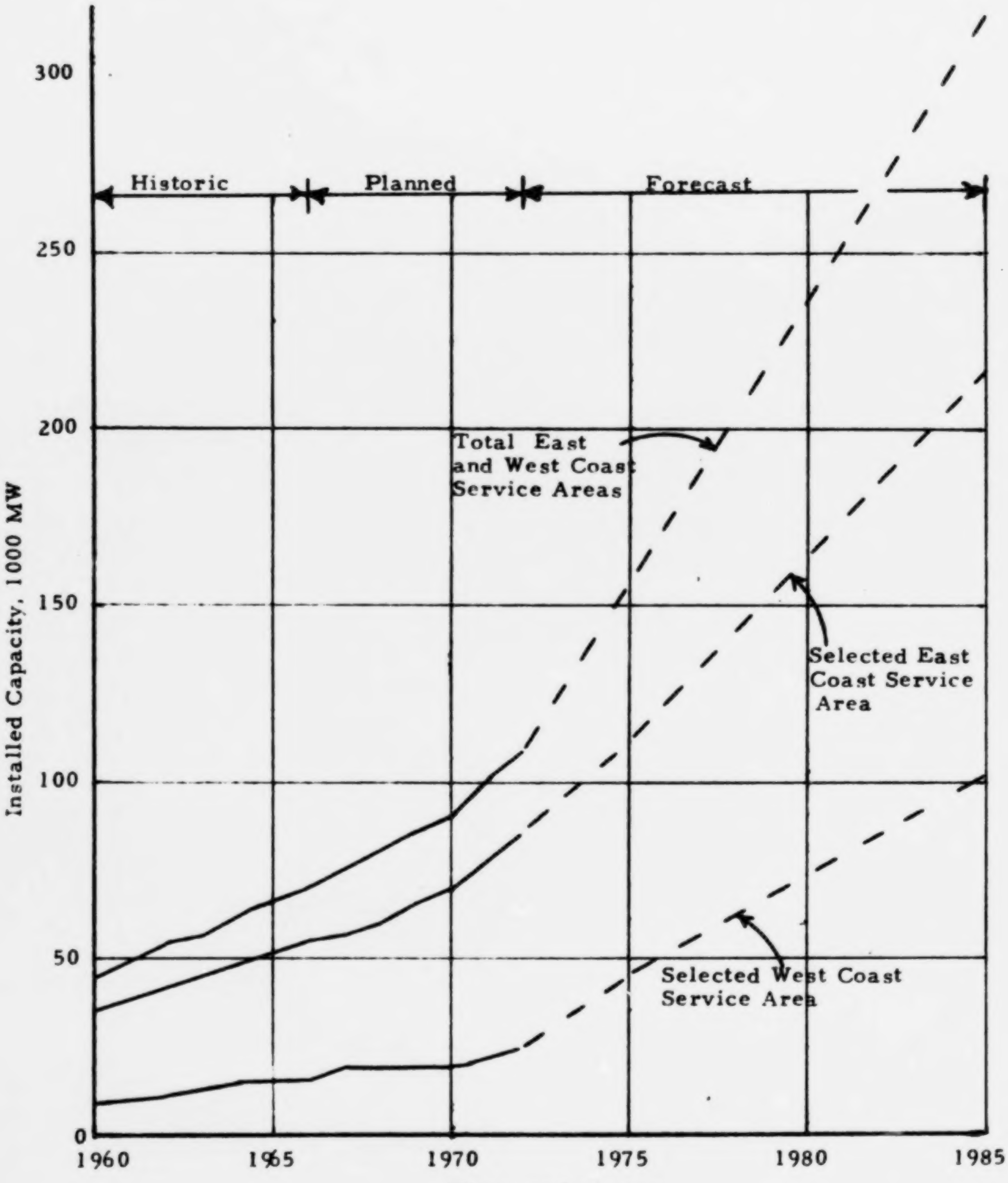

Figure 3.1.-1. GENERATION CAPACITY GROWTH IN SELECTED AREAS* ALONG EAST AND WEST COASTS (Excluding Hydroelectric)

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* See Tables 3.1.-1 - 3.1.-8.

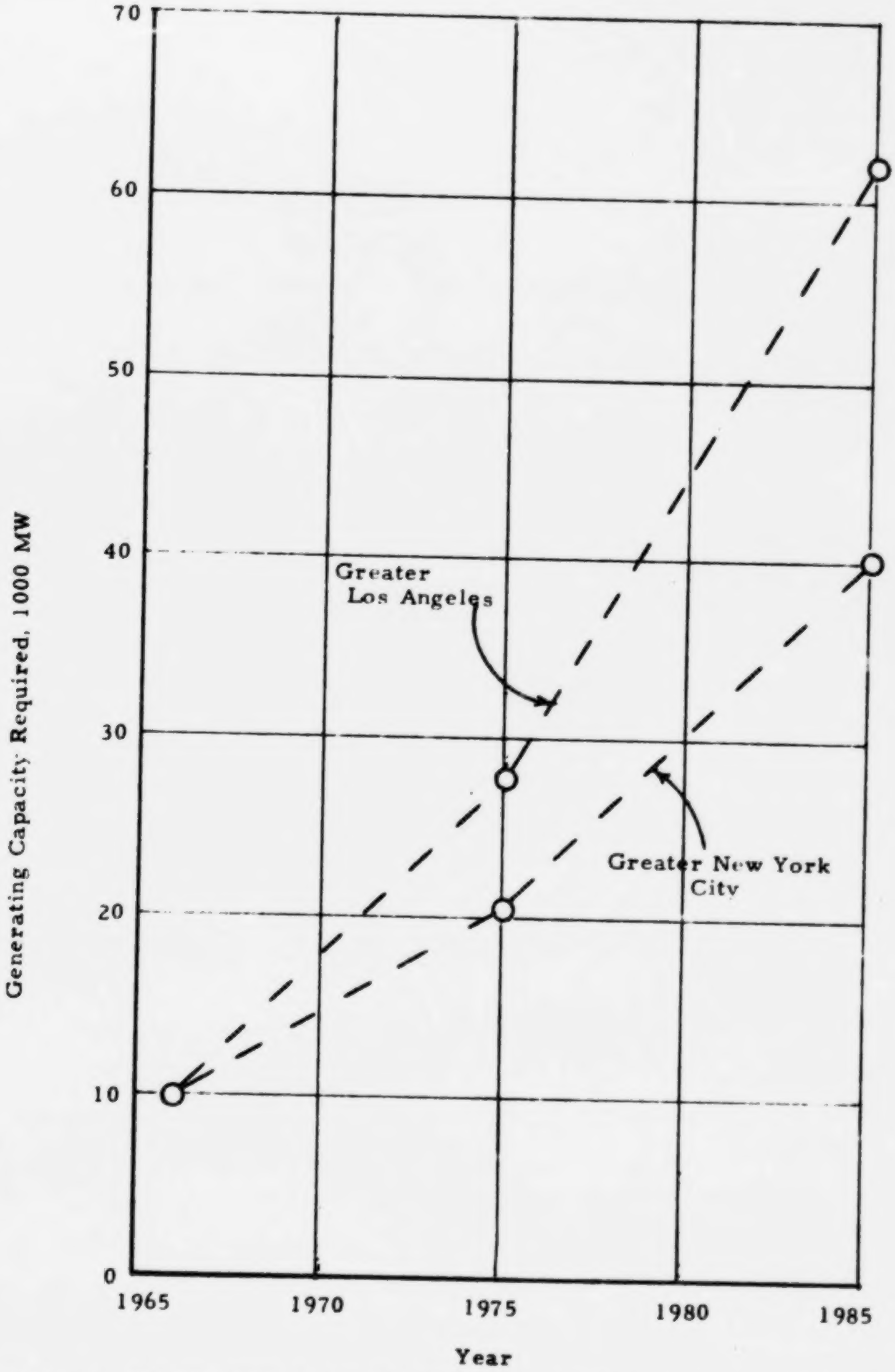

Figure 3.1.-2. PROJECTED GROWTH OF A 10,000-MW DEMAND BLOCK IN SELECTED METROPOLITAN AREAS[6]

projections presented assume only evolutionary, not revolutionary, developments. It is assumed that cities such as Los Angeles will continue to "import" large amounts of electric energy from outside their areas or states. Although a high proportion of the "imports" in past years has been generated in hydroelectric facilities, future imports will, of necessity, be generated to an increasing extent in fossil-fueled plants that are located at mine mouth or other locations outside of the metropolitan "air shed." Thus, future growth patterns have been assumed to follow traditional lines except for even more emphasis than in the past on keeping fossil-fueled stations in remote locations. This not only incurs abnormally high energy transportation costs, but is also inconsistent with our basic assumption that LNG can be burned in metropolitan areas without making a large contribution to air pollution. Adjustments are necessary to account for the discontinuities in past trends that will occur where LNG becomes the predominant fuel considered acceptable for use in metropolitan areas.

It would be difficult and probably unrewarding at this point to estimate precisely the amount of specific fuel displacement in 1975-1985 that will come about as a result of air pollution considerations. Adequate estimates for present purposes can be made, however, by making assumptions as to the number of 10,000-MW metropolitan blocks of steam-electric power to be in existence on each coast at the beginning and end of the period. This is done in Table 3.1.-11, which, for perspective, summarizes also the previously discussed regional projections based largely on the extrapolation of evolutionary trends.

The metropolitan blocks* for 1975 include two in the New York City area and one in the Philadelphia-Camden area on the East Coast, and two in the Los Angeles-San Bernardino area and one in the San Francisco area on the West Coast. These projections are based on traditional demand growth trends without assuming major new fuel displacement impact due to air pollution restrictions. By 1985, the three East Coast blocks will have increased to seven blocks, and three new ones will have been added,

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* Metropolitan blocks chosen for planning purposes to minimize redistribution of imported LNG.

Table 3.1-11. SUMMARY OF STEAM-ELECTRIC GENERATING CAPACITY PROJECTIONS

	1966	1975	1985
		Megawatts	
East Coast	53,370	110,000	215,000
West Coast	16,157	44,400	100,000
		Multiples of 10,000 MW	
East Coast	5.3	11.0	21.5
West Coast	1.6	4.4	10.0
		Estimated Number of 10,000-MW Metropolitan Demand Blocks*	
East Coast	1	3	10
West Coast	0	3	9

* Assumptions:

1. Nuclear power will continue to be unacceptable at metropolitan sites.
2. Coal and oil will become unacceptable at metropolitan sites, if not already so.
3. NG/LNG will be acceptable at all sites.
4. Locally sited, gas-fueled generating stations will displace an indefinite but substantial amount of capacity now (or planned to be) imported from outside the metropolitan areas.

including single blocks in the Boston area by 1983, in the Miami area by 1984, and in either the Tampa or Jacksonville area by 1985. On the West Coast, in addition to increasing the three 1975 blocks to seven, an eighth will be added in the San Diego area by 1985 and a ninth at a Pacific Northwest location by 1985.

Since even these 1985 projections may be derived by simply applying regional growth patterns to metropolitan growth, it is possible that higher numbers of blocks will actually develop. However, the substitution of metropolitan-sited LNG burning plants for remote coal- or oil-burning plants and additional hydroelectric facilities will, realistically viewed, be countered to some extent by the growth of nuclear power plants that are remotely located for safety reasons.

The necessary steps in the transition from the metropolitan fuel proportions as they exist in late 1974 to predominantly LNG fuel envisioned to start in 1975 must be developed in considerable detail to ensure the required coordination of supply arrangements and the phasing-out of older, less desirable arrangements. For example, there is some minimum size of generation plant in a given metropolis that will justify the importation of LNG for its use alone. A variety of plans might be considered, but we must make simplifying assumptions at this early stage of the study involving the economics of plant capital and operating costs.

As a basis for analyzing generation plant cost factors, we assume that the 10,000-MW metropolitan block of late 1974 will be increased by 1000 MW of LNG-fueled generation capacity per year during the 1975-1985 period, thus doubling its output capacity during that decade. This obviously presumes a high percentage rate of growth during the early part of the decade and a slow percentage rate of growth during the latter part. Such a pattern would not be expected from a trending of traditional growth patterns, but could turn out to be quite realistic in view of the strong incentives by 1975 1) to retire the then-existing pollution-producing plants, 2) to substitute LNG fuel for existing fuel in others, and 3) to create at a rapid rate a sufficient minimum fuel demand base to support a sizable LNG importation system.

3. 1. 1. East Coast (N. Y. C. -N. J.)

Figure 3. 1. -2 indicates that a 10, 000-MW block of electrical demand selected from the metropolitan New York City (N. Y. C. -J. J.) area in 1966 would be expected to grow to 20, 600 MW in 1975 and to 39, 900 MW in 1985. One could, of course, project the fuel requirements schedule of this particular 10, 000-MW block through the next few decades, or it would be possible to project the growth in demand and fuel requirements within a geographic area having a 10, 000-MW demand in 1966. As a measure of the opportunity for LNG to penetrate this metropolitan market as a replacement for less desirable fuels, it will be best to first examine the total metropolitan power generation requirement and project its growth. We have done this, assuming growth rates corresponding to the above figures and assuming no change in area import/export ratios.

As a basis for metropolitan area statistics, reference was made to Federal Power Commission Map M-78 showing "Principal Electric Facilities in the United States – 1966."[3] This map identifies 26 principal generating stations in the Greater New York City area, with a total generating capacity of 12,845 MW in 1966. When the growth of this capacity is scaled to the same projections as Figure 3. 1. -2, the 1975 capacity is projected as 26, 450 MW, and the 1985 capacity as 51, 300 MW. The year-by-year projections are given in Table 3. 1. -12 and Figure 3. 1. -3. These exhibits also give the corresponding fuel requirements based on the plant use factors and net heat rates indicated.

Estimates of average heat rates are based on a "conventional mix" of the various fuels. Gas-fueled plants would probably have above average heat rates, but efficiency gains through the use of LNG cold might, in part, offset these losses. In any event, the stated averages are sufficiently accurate for present purposes.

Table 3.1.-12. EAST COAST METROPOLITAN CHARACTERISTICS DATA (N.Y.C.-N.J.)

Year	Installed Capacity, 1000 MW	Plant Use Factor	Plant Output, 10^9 kWhr	Net Heat Rate, Btu/kWhr	Fuel Input, 10^{12} Btu
1975	26.450	0.50	115.8	9400	1088
1976	28.935	0.497	126.0	9370	1181
1977	31.420	0.494	136.0	9340	1270
1978	33.905	0.491	145.6	9310	1356
1979	36.390	0.488	155.3	9280	1442
1980	38.875	0.485	165.2	9250	1528
1981	41.360	0.482	174.3	9220	1608
1982	43.845	0.479	183.9	9190	1690
1983	46.330	0.476	193.0	9160	1769
1984	48.815	0.473	202.0	9130	1845
1985	51.300	0.470	211.0	9100	1920

Since the growth of electrical demand is a gradual rather than stepwise phenomenon, it would be unrealistic to assume the sudden addition of 10,000 MW of generating capacity to an existing 26,450-MW system (unless considerable exporting of energy were contemplated), even though the addition of such large blocks will not seem as spectacular in 1975 as it does today. In any event, we are primarily concerned with the requirements for LNG fueling facilities. It appears that there is ample opportunity (as well as justification) for the displacement of less desirable fuels in the N.Y.C.-N.J. area to permit the addition (partly by substitution) of 10,000-MW increments of LNG-fueled capacity in 1975 and thereafter.

Figure 3.1.-4 shows the projected growth in number of 10,000-MW power blocks on the East Coast. There will be 3 in 1975 and 10 in 1985, with approximate dates and locations of additions scheduled as indicated.

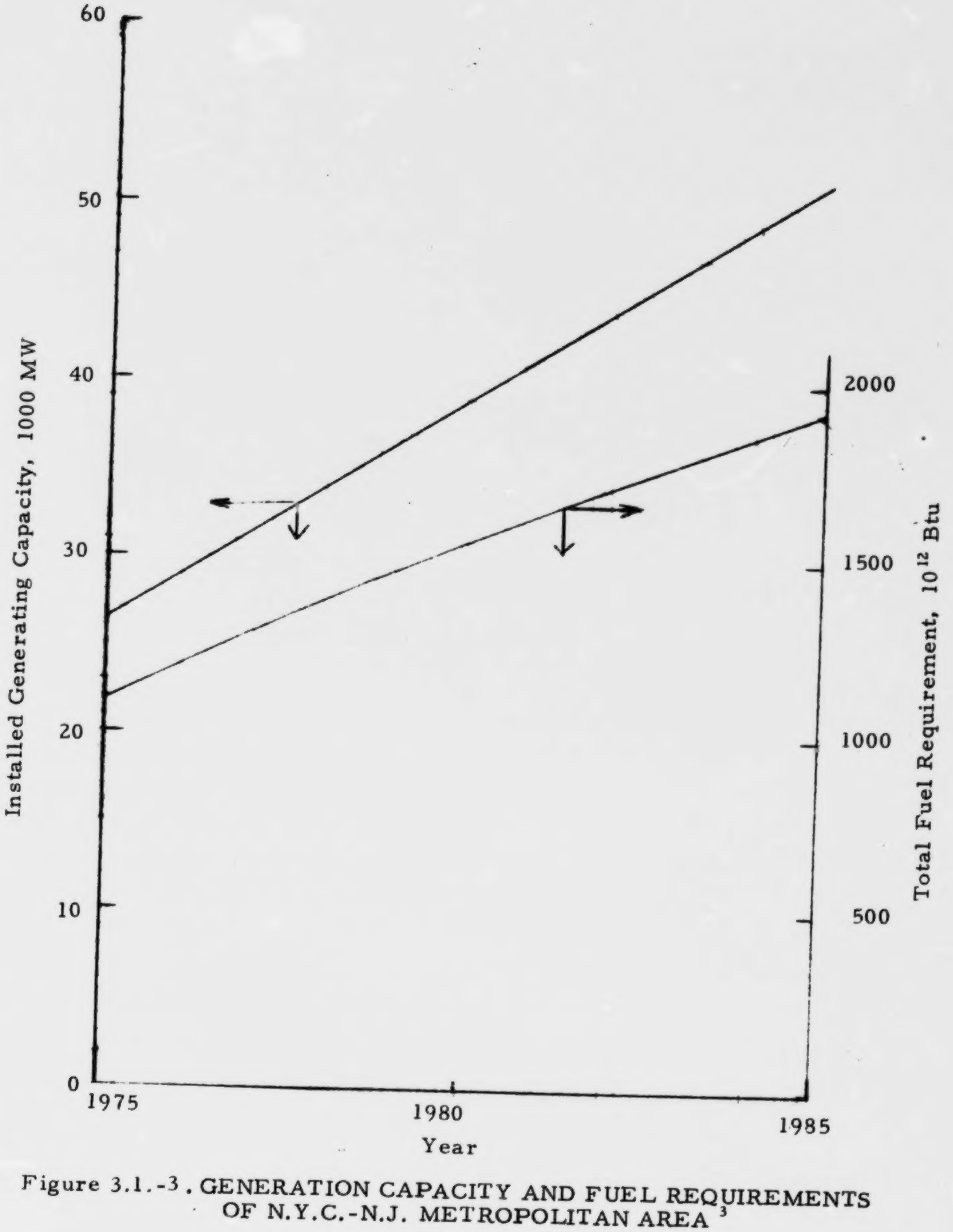

Figure 3.1.-3. GENERATION CAPACITY AND FUEL REQUIREMENTS OF N.Y.C.-N.J. METROPOLITAN AREA [3]

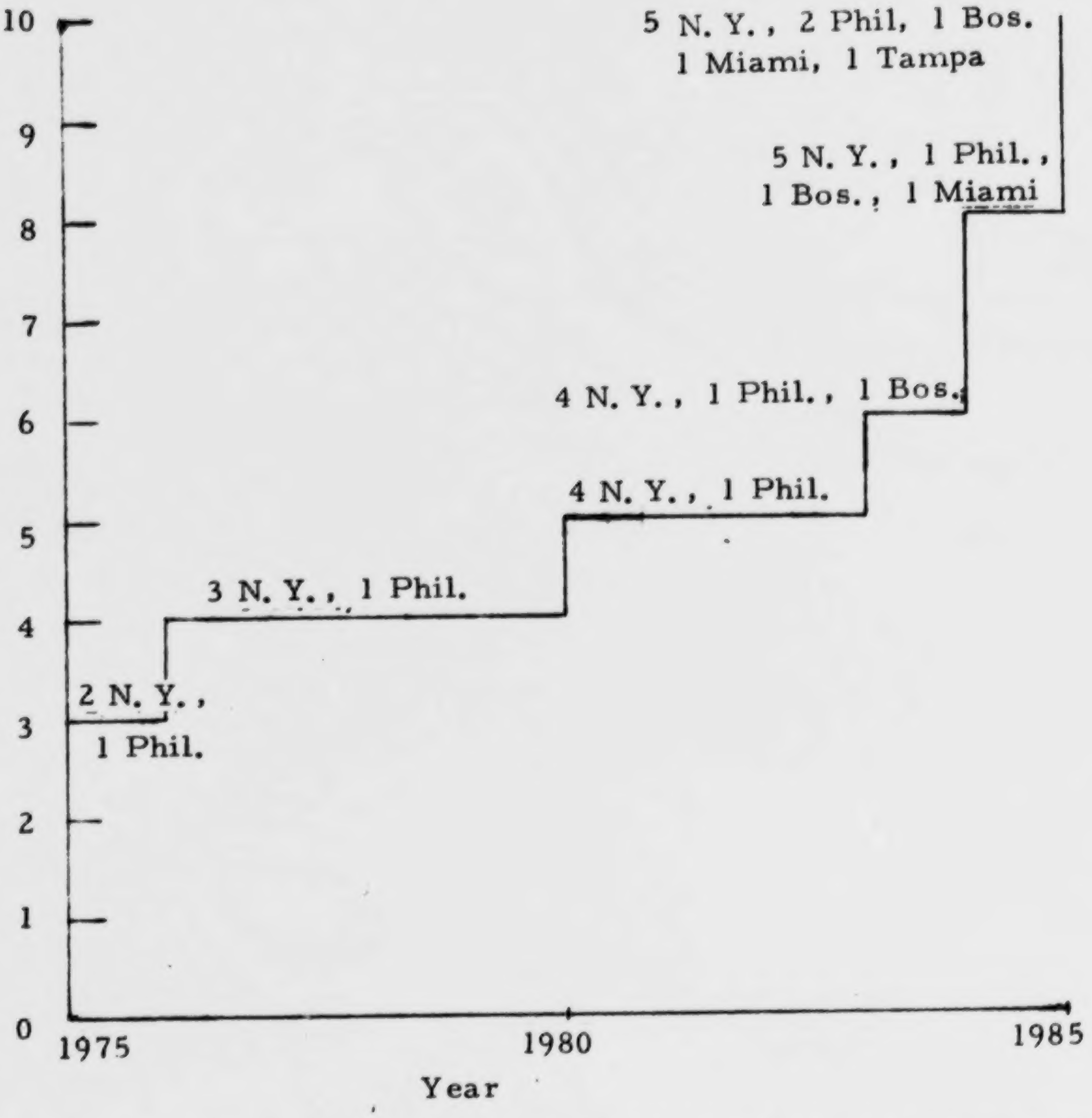

Figure 3.1.-4. PROJECTED NUMBER OF 10,000-MW METROPOLITAN POWER BLOCKS ON EAST COAST

3.1.2. West Coast (Los Angeles)

The same procedures were followed in projecting Metropolitan Los Angeles requirements and those of other West Coast centers, as shown in Table 3.1.13, Figure 3.1.-5, and Figure 3.1.-6.

Even though Los Angeles was behind New York City in 1966 installed generating capacity, it is expected to have more than New York in 1975 and to continue growing at a more rapid rate through 1985. By 1985, the West Coast will have nine blocks of 10,000 MW or more, including five in the Los Angeles area.

Table 3.1.-13.WEST COAST METROPOLITAN CHARACTERISTIC DATA

Year	Installed Capacity, 1000 MW	Plant Use Factor	Plant Output 10^9 kWhr	Net Heat Rate, Btu/kWhr	Fuel Input, 10^{12} Btu
1975	25.850	0.500	113.3	9400	1064
1976	29.085	0.497	126.6	9370	1187
1977	32.320	0.494	139.6	9340	1303
1978	35.555	0.491	152.7	9310	1421
1979	38.790	0.488	165.7	9280	1538
1980	42.025	0.485	178.4	9250	1650
1981	45.260	0.482	190.9	9220	1759
1982	48.495	0.479	203.2	9190	1870
1983	51.730	0.476	215.7	9160	1975
1984	54.965	0.473	227.5	9130	2075
1985	58.200	0.470	239.5	9100	2180

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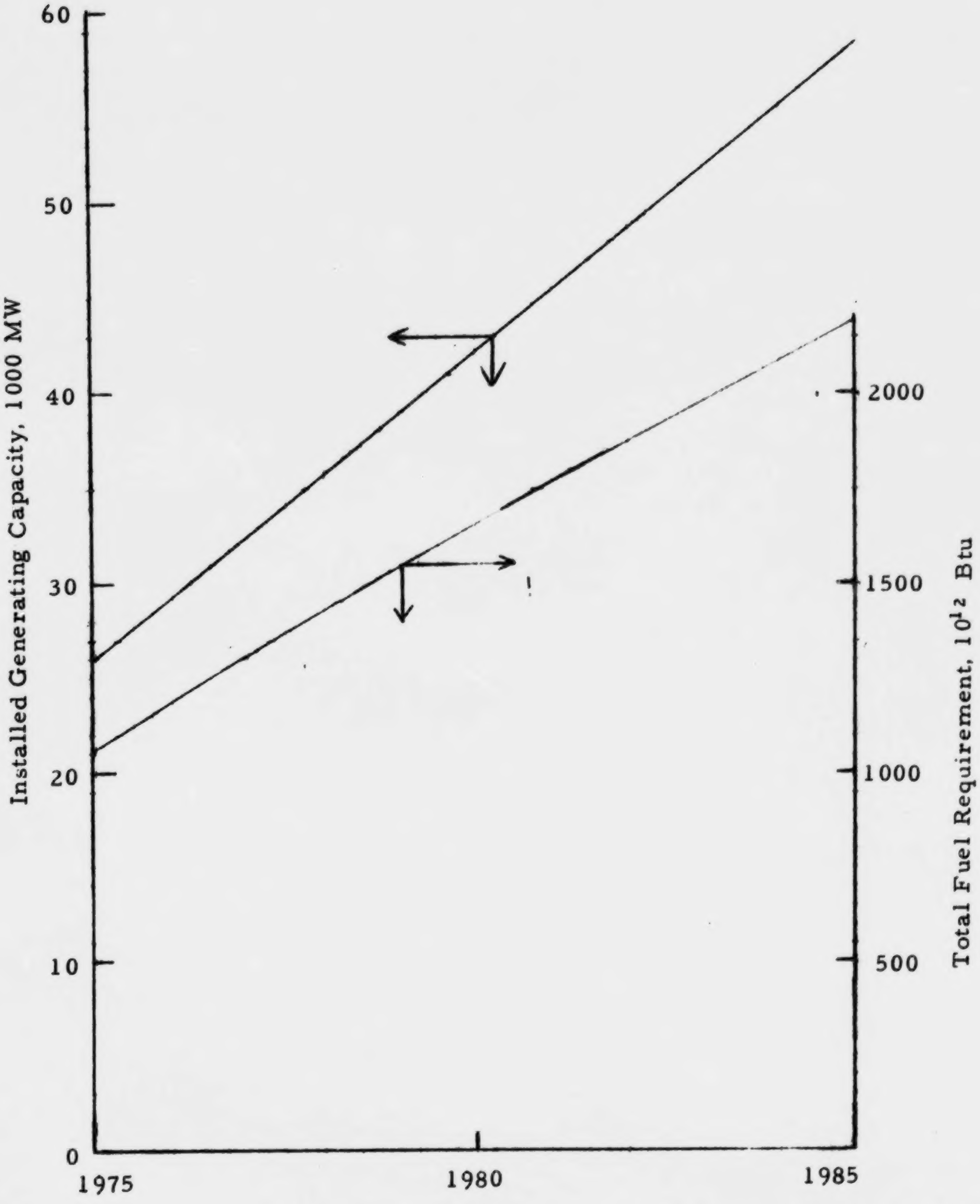

Figure 3.1.-5. GENERATION CAPACITY AND FUEL REQUIREMENTS OF LOS ANGELES METROPOLITAN AREA

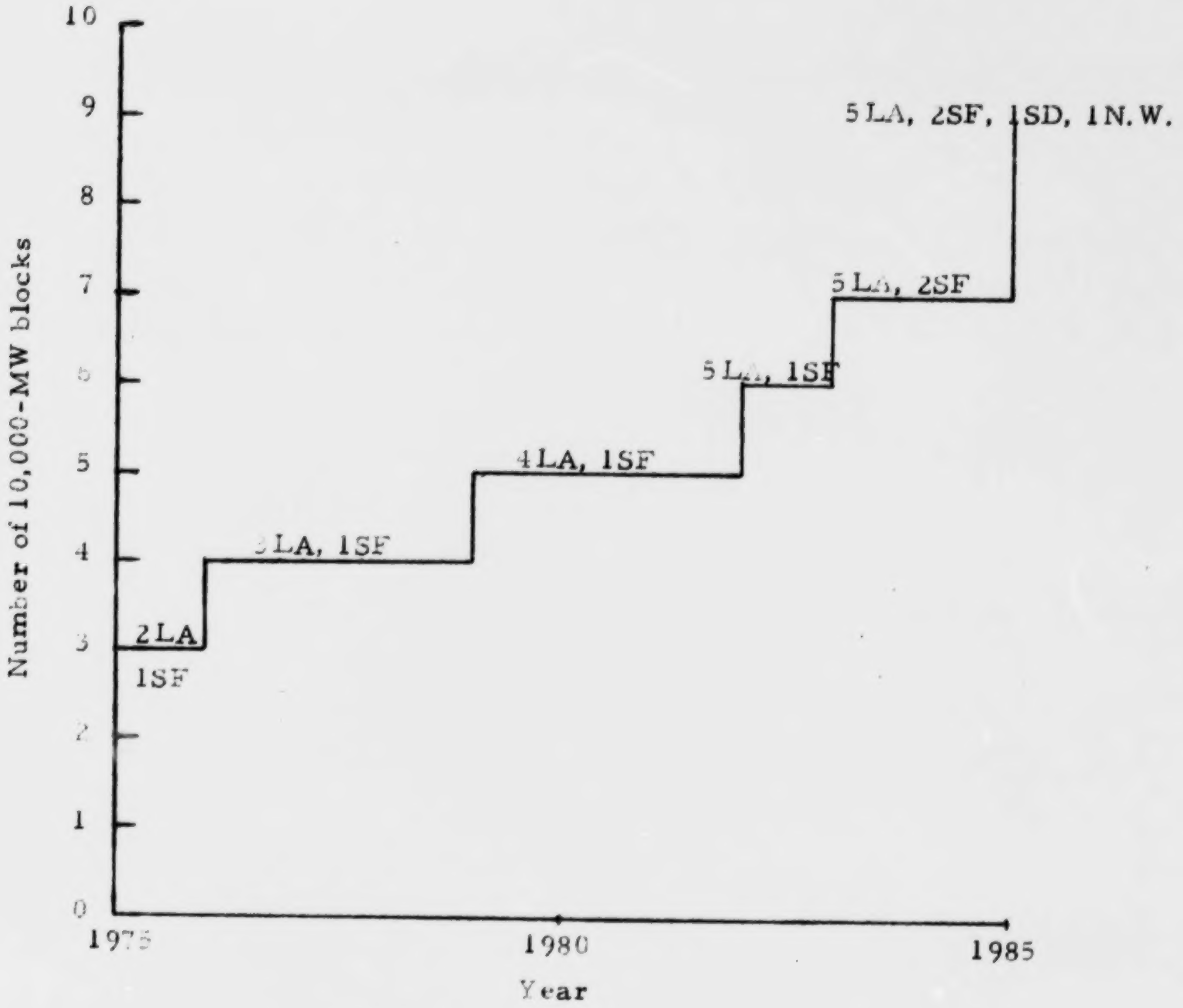

Figure 3.1-6. PROJECTED NUMBER OF 10,00-MW METROPOLITAN POWER BLOCKS ON WEST COAST

3.2. Power Generation

In this section the specific advantages of gas-fired electric power generating plants over coal and oil plants are discussed and quantitatively evaluated. Section 3.2.1., Power Plant Description, outlines the flexibility of natural gas fuel in affording performance improvements in the conventional (Rankine) process cycle. It also qualitatively evaluates the advantages of natural gas plants in such areas as fuel-handling facilities, steam generator size, stack height, and personnel requirements.

Sections 3.2.2., Investment Costs, and 3.2.3., Operating Costs, provide a cost comparison between natural gas, oil, and coal power plants for locations on both the Eastern and Western Seaboards. Section 3.2.4. gives the investment and operating costs for vaporized LNG. This then provides the basis for a later comparison of LNG-fueled electrical generation and the ultimate comparison of bus bar energy costs. All the generating costs are based upon a 10,000-MW plant base-loaded at a 90% annual load factor. The electrical generating cost figures for each alternative are estimated for the years 1970, 1975, 1980, or 1985 under the assumption that the 10,000-MW plant first goes into operation in one of these years.

Section 3.2.4. provides a detailed breakdown of the costs of imported LNG for a 10,000-MW electrical plant on both Seaboards. The major cost components covered are as follows:

1. LNG cost after liquefaction
2. Transportation costs
3. Storage costs
4. Vaporization costs

The summation of these cost components provides a fuel cost for imported LNG which is directly comparable to the fuel costs of coal, oil, and natural gas in Section 3.2.3., Operating Costs.

The imported LNG cost facilitates a final comparison of bus bar electrical generation costs for power plants designed for either LNG, natural gas, oil, or coal. This comparison is presented in Section 3.2.4.7, Power Costs. For the oil and coal alternatives, however, additional plant investments and operating costs have been assumed for flue gas desulfurizing. The additional investment cost for desulfurizing was assumed to be \$15/kW; the additional operating cost was assumed to be \$0.10/million Btu of fuel burned.*

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* Basis supplied by NAPCA.

3.2.1. Power Plant Description

Whether they burn coal, oil, or gas, essentially all central thermal power stations in the U.S. use the same (Rankine) process cycle. Since this practice is expected to continue into the foreseeable future, the major findings of this report are based on this premise. It should not be inferred, however, that LNG usage is limited to conventional cycles. In fact, the opposite is true. LNG is a premium-quality fuel for, for example, the gas turbines currently being installed at an accelerated rate to meet urgent standby, peakshaving, and end-of-line load balancing requirements of the major electric utilities. Gas turbine generating capacities are continually being increased to enable them to fulfill an ever wider range of applications.

For example, fluid injection at the combustor is one method of increasing gas turbine generating station capacity while reducing unit costs. At the 1965 American Power Conference several variations of this cycle were described by several Westinghouse Electric Corp. representatives in a paper entitled, "Future Role of Gas Turbines in Power Generation."[4] The so-called combined cycle also presents possibilities for the beneficial exploitation of clean fuels such as LNG. In the simplest application, where the gas turbine exhaust is used as heated combustion air for a conventional steam generator, the advantages demonstrated to date have been minimal for various reasons, but all possibilities have probably not been exhausted. A more attractive possibility, though one requiring considerably more design refinement, is a proposed cycle in which a pressurized steam generator receives its combustion air directly from the gas turbine compressor, with the gas power turbine being driven by steam generator exhaust gases. In such a design, a clean-burning fuel such as natural gas is necessary to minimize fouling and erosion of the tightly compacted heat exchange surfaces.

Returning again to the conventional Rankine cycle systems, the type of fuel burned has a significant influence on the design, capital cost, and operating costs associated with a central power station. Some of the major influences are qualitatively reviewed below.

3.2.1.1. Fuel-Handling Facilities

In the large metropolitan centers such as New York or Los Angeles the quality of pipeline natural gas is highly uniform. Only rarely is it necessary for an electric utility to do any significant amount of gas conditioning before burning this fuel. Some pressure regulation by the user may be necessary to meet his individual requirements, but, generally speaking, his capital investment in gas-handling facilities is ordinarily negligible.

Where oil is burned, storage tanks are necessary. This requires capital expenditures for dock facilities, tankage, real estate, safety provisions, and handling equipment (pumps, piping, heaters, gaging equipment, etc.). Oil quality is highly variable depending on its geographic origin, degree of pretreating, sulfur and vanadium content, etc. This creates a wide range of complex, and often costly, operating problems.

Where coal is burned, the necessary storage, cleaning, and handling equipment is highly visible. Control laboratories must be maintained to continuously check coal composition and storage temperature (to avoid spontaneous combustion). Also procurement-utilization scheduling is a constant problem. In recent years, the effort to reduce delivered coal costs via unit trains has imposed a heavy burden on terminal facilities at the receiving end as well as at the mines. Thus, there has been a significant increase in central station costs for this reason alone. Not to be overlooked, of course, is the high cost of disposing of ash in an acceptable manner.

3.2.1.2. Generator Size

For a given generating capacity, a coal-fired steam generator is typically much larger than a gas-fired furnace. In many ways, this increases capital costs. Keeping heat exchange surfaces clean in a coal furnace requires the installation and operation of extensive facilities (such as soot blowers), all of which show up in overall capital and operating costs.

3.2.1.3. Stack Height

Coal-fired units are being equipped with smokestacks of heights approaching 1200 feet that cost millions of dollars. This is necessary to diffuse noxious gases, such as sulfur dioxide, and residual particulate matter over wide expanses of air volume and land area, thereby diluting their harmful effects and avoiding highly visible local nuisances. Gas-fired units have much shorter, less-costly stacks. The size and cost of breeching from the furnace to the stack are also much greater in coal-fired than in gas-fired plants. This is especially significant where fly-ash precipitators or other processing equipment are inserted between the furnace and the stack.

3.2.1.4. Personnel Requirements

Many operations in coal-fired and, to some extent, oil-fired units require personnel for which there is either no counterpart in gas-fired units or for which the need is greatly reduced. The most obvious savings are in fuel-handling operations. Maintenance needs are also much less severe where clean, sulfur-free fuels such as natural gas or LNG are used.

The operating personnel requirements are less in gas-fired plants. For example, there are no problems of clinker formations interfering with ash and slag removal and causing local overheating of furnace parts. Ash handling is also eliminated. Soot-blowing and tube-cleaning operations, in general, are eliminated or greatly reduced. Corrosion and erosion problems are minimized, and electrostatic precipitators are not required for fly-ash removal. Smoke problems are essentially eliminated. All of these operations require manpower with its supporting services and overhead charges. Also, where manpower has been replaced by automation, the cost of gas-fired plant automation is usually much lower than that in coal- or oil-fired plants.

3.2.2. Investment Costs

Investment costs for gas, oil, and coal electric generating plants have been established on a $/kW basis for 10,000-MW plants. These costs reflect current technology and construction costs for both the Eastern and Western Seaboards. Therefore, the investment costs show a sizable difference between gas, oil, and coal plants for the reasons cited in Section 3.2.1., Power Plant Description. In addition, the power plant costs on the Eastern Seaboard are higher than those on the Western, primarily because of the enclosed construction required there. The Western Seaboard climatic conditions often allow semioutdoor or outdoor construction.[5]

In all cases, these investment costs are total cost estimates and cover the costs of developing a new plant. They are, therefore, slightly higher than what would be required for adding to an existing plant site. Table 3.2.-1. provides an investment cost summary which includes projected costs for 5-year increments to 1985.

Table 3.2.-1. INVESTMENT COSTS FOR NEW 10,000-MW ELECTRIC GENERATING PLANTS
(Not Including Desulfurizing Equipment)

Year	Gas	Oil	Coal
		$/kW	
		Eastern Seaboard Costs	
1970	95	120	140
1975	109	137	160
1980	125	157	183
1985	143	180	210
		Western Seaboard Costs	
1970	85	110	130
1975	98	127	150
1980	113	147	173
1985	130	170	200

These cost estimates do not include the cost of desulfurizing equipment.

The investment cost projections for years 1975, 1980, and 1985 in Table 3.2.-1. are based upon historical electric light and power construction cost trends. Figure 3.2.-1. presents the Handy-Whitman Construction Cost Index which was used to establish the construction cost increase for each coastal area. The North Atlantic construction cost growth rate was 2.74%/yr; the Pacific rate was 2.88%/yr. Compounded for 5-year increments the respective values are 14.5 and 15.3% increases.

The investment costs for the base year 1970 include a $25/kW additional cost for the oil-fueled plant over the gas-fueled plant and $45/kW additional cost for the coal-fueled plant over the gas plant. These cost differences are all before additional desulfurizing equipment is considered. Of the $45/kW added cost for coal plants over gas, $13-$14/kW can be attributed to the coal- and ash-handling facilities. An amount estimated at $25/kW is attributed to the additional boiler, breaching, and soot-blowing equipment required by coal plants. The additional differential of $6-$7/kW is charged to extra land requirements, taller stacks, and added maintenance and laboratory facilities.

The annual fixed charges per kWhr attributed to investment cost are obtained by applying the 14% approximate per year per dollar of original investment rate advocated by the Atomic Energy Commission for investor-owned power stations. In addition, it is also necessary to consider the annual plant load factor, which, for this study, is assumed to be 90%. The fixed charge calculation is made as follows:

$$\text{Fixed charge (mills/kWhr)} = \frac{\text{Invest. Cost (\$/kW)} \times 0.14\ /\text{yr}}{\text{Load Factor} \times \text{hr/yr}}$$

$$= \frac{\text{Invest. Cost (\$/kW)} \times 0.14}{0.90 \times 8760}$$

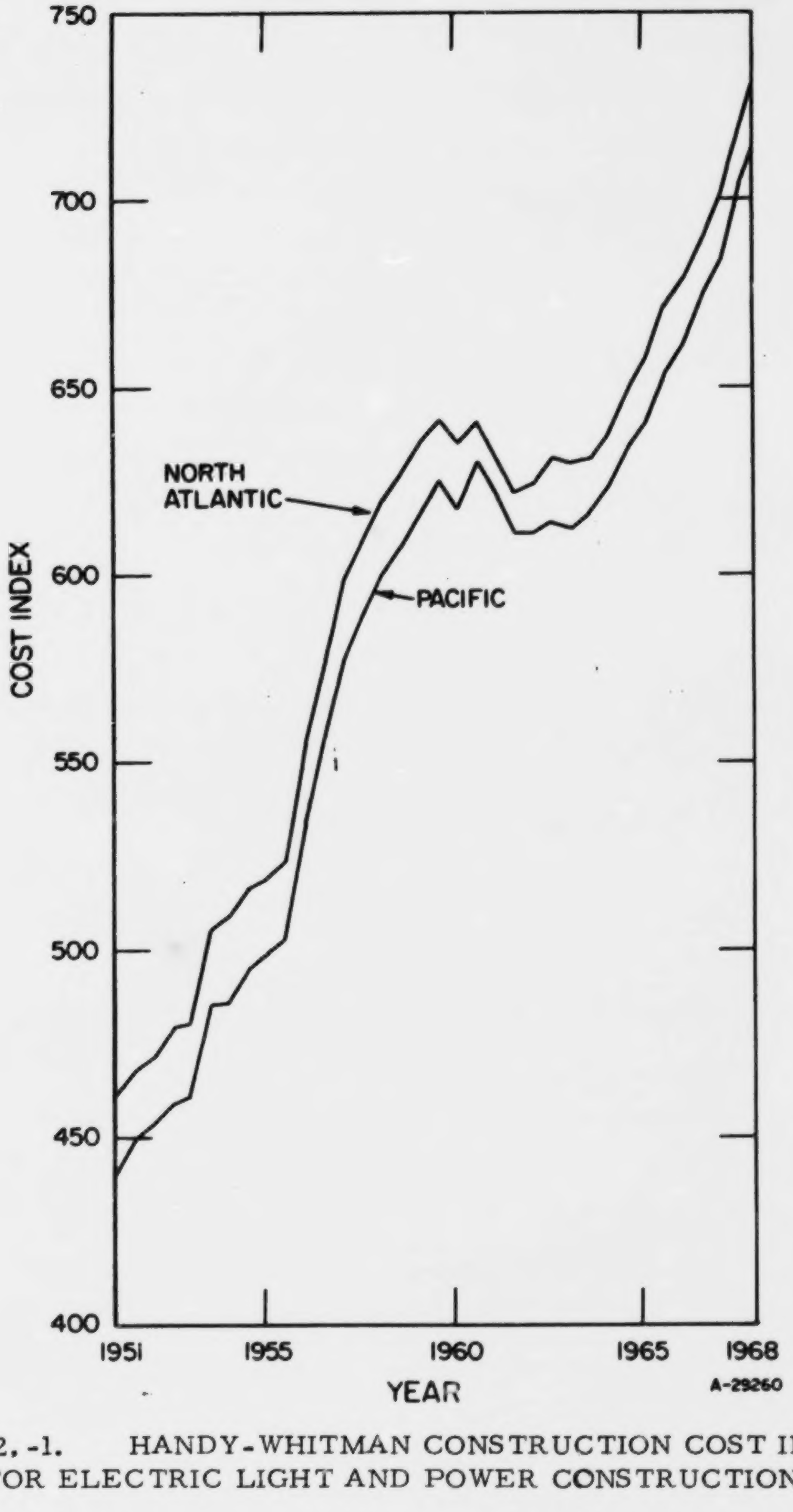


Figure 3.2.-1. HANDY-WHITMAN CONSTRUCTION COST INDEX FOR ELECTRIC LIGHT AND POWER CONSTRUCTION

The fixed charges that result from applying the above formula to the investment costs given in Table 3. 2. - 1. and an additional $15/kW for desulfurizing equipment are presented in Table 3. 2. -2. These costs reflect the expected fixed charge assuming the plant first goes into operation on the given year 1970, 1975, 1980, or 1985.

Table 3. 2. -2. FIXED CHARGES FOR NEW 10,000-MW ELECTRIC GENERATING PLANTS (Includes $15/kW Investment Cost for Desulfurizing Equipment for the Oil and Coal Alternatives)

Year	Gas	Oil	Coal
	mills/kWhr		
Eastern Seaboard Costs			
1970	1.69	2.40	2.76
1975	1.94	2.71	3.11
1980	2.22	3.06	3.52
1985	2.54	3.47	4.00
Western Seaboard Costs			
1970	1.51	2.22	2.58
1975	1.74	2.53	2.93
1980	2.00	2.88	3.34
1985	2.31	3.29	3.82

The assumed $15/kW investment cost for desulfurizing equipment has not been escalated as have the plant costs. It is further assumed that future technological advances will help maintain this position of the cost at a constant $15/kW. This assumption is perhaps too optimistic and may unduly favor the oil and coal alternatives. Under the assumption of escalating investment costs for desulfurizing equipment, the 1985 cost would be 50% higher. The assumed high load factor also favors oil- or coal-fired plants.

3.2.3. Operating Costs

Operating costs that are fuel plus labor, supplies, etc., constitute the major portion of the bus bar electric costs. They can amount to more than double the annual fixed charges for any given installation. Therefore, they represent a prime area of concern in all electric generation economic studies.

The operating costs are broken down into two major categories: fuel costs and all other operating costs. The other costs include such items as maintenance and repairs, wages and supervision, and water, lubrication and supplies. The fuel cost, however, is the major cost item in the operating cost segment and can amount to 80-90% of the total operating costs.

The fuel cost component is established by multiplying the net heat rate for each type of plant by the respective fuel cost on a Btu basis. The appropriate net heat rates for 1970 plant start-up are as follows:

Plant	1970 Net Heat Rates, Btu/kWhr
Gas	9400
Oil	9250
Coal	8900

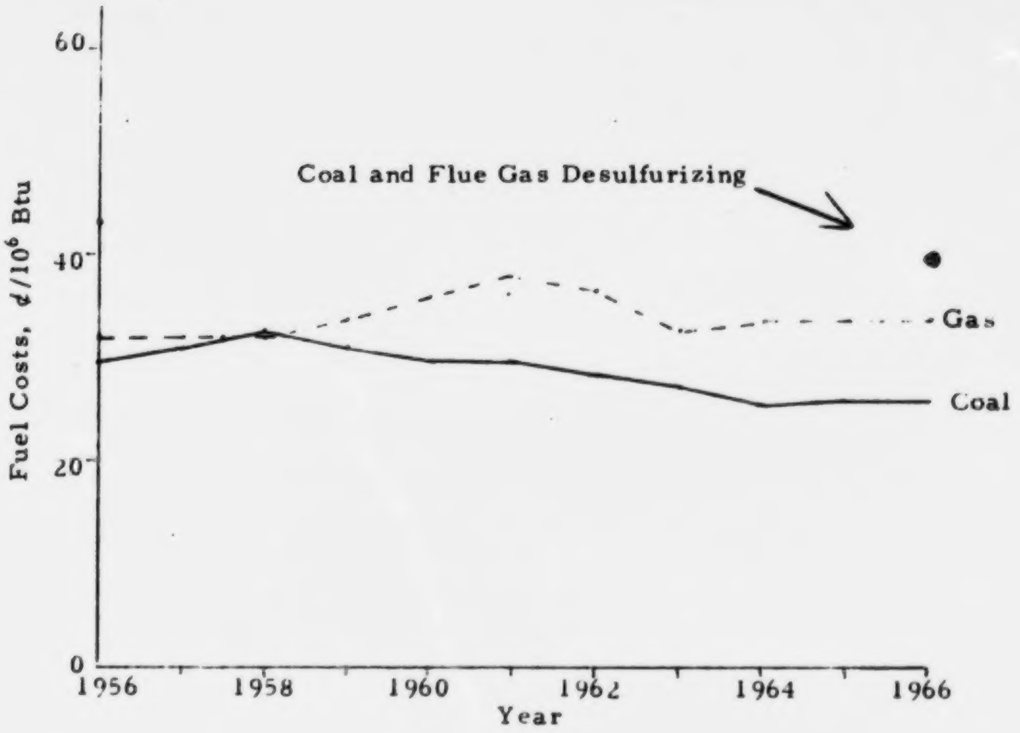
To determine the projected fuel costs for the years following 1970, it was necessary to consider the trend toward heat rate improvements in new plants. An improvement in heat rate of approximately 1%/year has been achieved over the past 10 years. However, considering theoretical limits and that there is no guarantee that past improvement rates could be maintained for the period 1970 to 1985, it has been assumed that an improvement of 2% per each 5-year increment can be achieved. These heat rate improvements are used in Table 3.2.-3, which is presented later following a discussion of the gas, oil, and coal fuel costs for both coastal areas.

The costs per million Btu for fuels burned in power plants on the East and West Coasts are plotted for the years 1956 through 1966 in Figures 3.2.-2 and 3.2.-3. The data were taken from a National Coal Association (NCA) publication, Steam-Electric Plant Factors – 1966. The Middle Atlantic region data are used for the East Coast installation and Pacific region data for the West Coast.

The one point on each figure (shown for the year 1966) shows what the equivalent coal or oil cost at each date would have been if its actual cost had been adjusted upward by the corresponding capital and operating costs of flue gas desulfurization had it been in use at the costs specified in the project work statement. The operating cost increment was 10¢/million Btu, as specified. The capital cost increment was computed based on the \$15/kW of installed capacity for flue gas desulfurizing equipment, as specified, taking into account the 90% plant load factor and the appropriate plant heat rate in each year, and assuming a fixed charge on invested capital of 14%/yr. More specifically, the capital cost increment was –

$$\frac{(0.14)\ (\$15/\text{kW})}{(8760\ \text{hr/yr})\ (\text{plant load factor})\ (\text{million Btu/kWhr heat rate})}$$

Thus, the equivalent coal and oil costs per unit of heating value required to produce stack gas conditions in oil- and coal-fired plants comparable to those in natural-gas-fueled plants without stack gas treatment are 12.9¢-13.0¢/million Btu higher than their respective first costs.



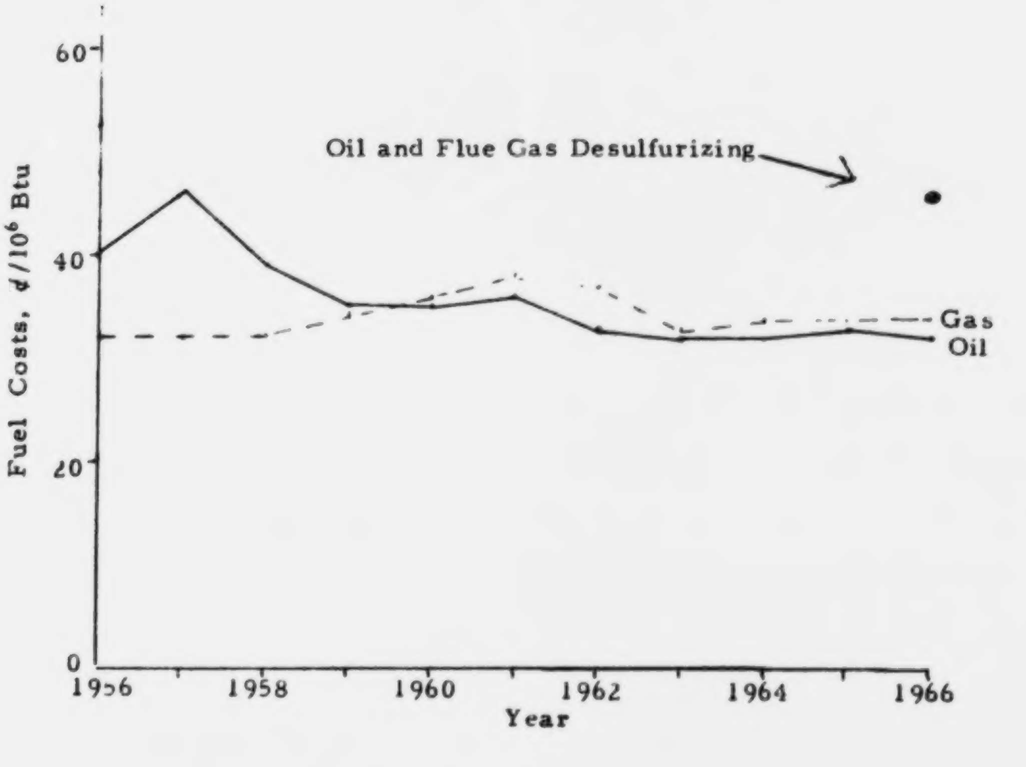


Figure 3.2.-2. MIDDLE ATLANTIC FUEL COST TRENDS IN STEAM ELECTRIC UTILITIES

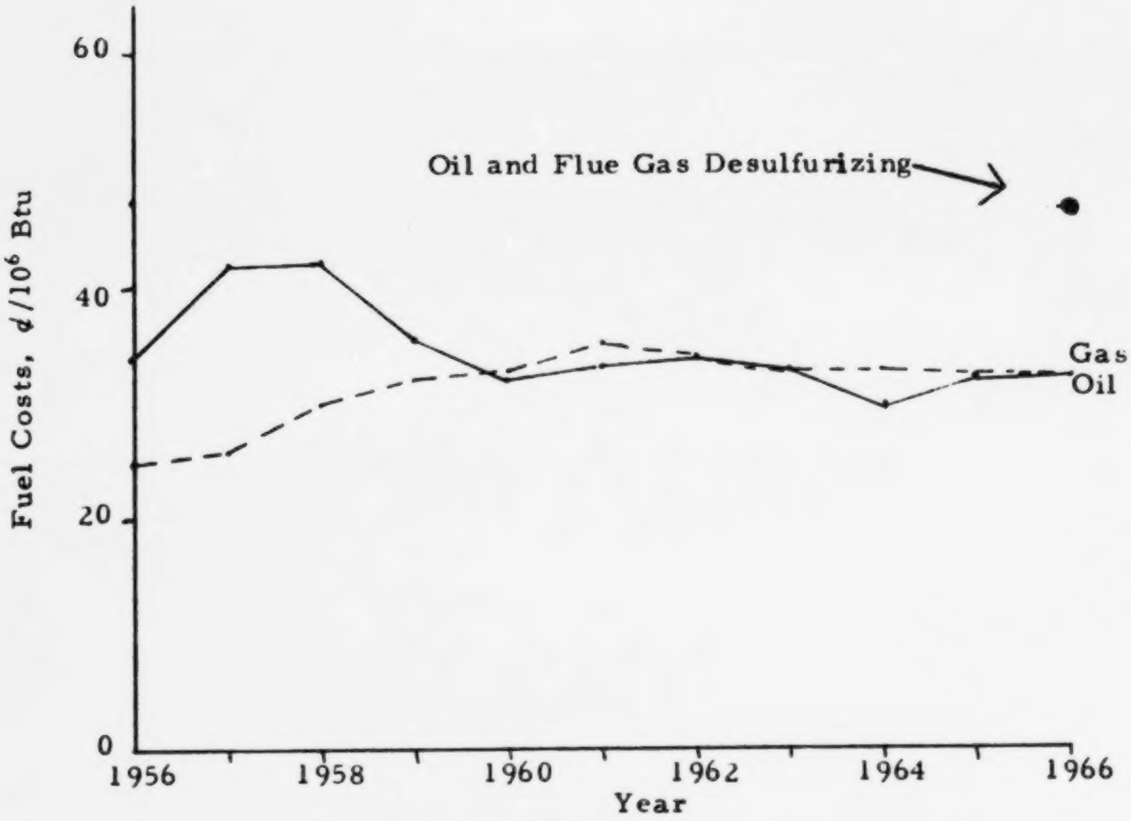


Figure 3.2.-3. PACIFIC FUEL COST TRENDS IN STEAM ELECTRIC UTILITIES

Few data are available on coal prices for the West Coast area power plants because, historically, coal has not been available at competitive prices for this market. Recent trends in coal slurry pipeline technology and the use of unit trains, however, could change this situation. Therefore, based upon such future possibilities, the price of coal delivered to the West Coast area could be as low as 30¢/million Btu. This figure is based upon an assumed mine mouth cost of 12¢/million Btu and an 18¢ million Btu delivery cost. Of course, both figures are highly speculative and are used here only for facilitating the evaluation of West Coast coal-fired electrical generating plants.[7,8]

Table 3.2.-3 provides the fuel costs expected on both a Btu and kWhr basis. In this summary, the fuel costs on a Btu basis are taken from historical averages in the National Coal Association publication and are assumed to remain constant over the projected period. Included in the oil and coal costs is the 10¢/million Btu cost for desulfurizing flue gases. Since the heat rates for each fuel are projected to decrease by 2% per a 5-year increment, the projected fuel cost per kWhr decreases.

Additional operating costs are presented in Table 3.2.-4. These costs are assumed to be equal for either an eastern or western plant location. Included in "other costs" are maintenance and repairs, wages and supervision, and water, lubrication, and supplies. These costs are estimated and totaled for 1970. For future years the cost has been escalated by 3% per year, or 15% for a 5-year time span.

The maintenance and repairs cost of 0.14 mill/kWhr for gas in base year 1970 is considerably less than the 0.23 mill/kWhr for both oil and coal. Part of this differential reflects the nearly complete absence of fuel- and ash-handling expense, which averages 0.04 mill/kWhr for coal-burning stations. The bulk of the savings, however, results from lower maintenance and repair costs for the boiler plant.[5] The wages and supervision component is the smallest. Gas is estimated at 0.08¢, oil at 0.09¢, and coal 0.10¢/kWhr. These relatively low figures reflect the economies of large steel plants in excess of 10,000 MW. The slightly higher values for coal and oil reflect the added manpower required for coal and ash handling. The water, lubrication, and supply cost is 0.20¢ for gas, 0.23¢ for oil, and 0.25¢ for coal. The differential cost reflects the added cost of supplies due to the additional plant facilities in oil and coal plants.

Table 3.2.-5 presents the summation of the fuel cost and other operating costs from Tables 3.2.-3. and 3.2.-4. The projected total operating cost shows surprisingly little inflation. The somewhat constant costs result from expected improvements in heat rates, which partially offset the effects of increasing labor and supply costs.

Table 3.2 -3. FUEL COSTS FOR 10,000-MW ELECTRICAL GENERATING PLANTS

(Includes 10¢/million Btu Operating Costs for Desulfurizing Oil and Coal Flue Gas)

Eastern Seaboard Costs

Year	Gas, 34¢/10^6 Btu	Oil, 32¢ + 10¢/10^6 Btu	Coal, 26¢ + 10¢/10^6 Btu
	mills/kWhr		
1970	3.19	3.88	3.20
1975	3.12	3.80	3.14
1980	3.06	3.73	3.08
1985	3.00	3.65	3.02

Western Seaboard Costs

Year	Gas, 32¢/10^6 Btu	Oil, 32¢ + 10¢/10^6 Btu	Coal, 30¢ + 10¢/10^6 Btu
	mills/kWhr		
1970	3.01	3.88	3.56
1975	2.94	3.80	3.49
1980	2.88	3.73	3.42
1985	2.83	3.65	3.35

Table 3 2.-4. OTHER OPERATING COSTS FOR 10,000-MW ELECTRICAL GENERATING PLANTS ON BOTH EASTERN AND WESTERN SEABOARDS

Year	Item	Gas	Oil	Coal
		mill/kWhr		
1970	Maintenance and Repairs	0.14	0.23	0.23
	Wages and Supervision	0.08	0.09	0.10
	Water, Lubrication, and Supplies	0.20	0.23	0.25
	Total	0.42	0.55	0.58
1975		0.48	0.63	0.67
1980		0.55	0.73	0.77
1985		0.63	0.83	0.88

Table 3.2.-5. TOTAL OPERATING COSTS FOR 10,000-MW ELECTRICAL GENERATING PLANTS

Year	Gas	Oil	Coal
	mills/kWhr		
Eastern Seaboard			
1970	3.61	4.43	3.78
1975	3.60	4.43	3.81
1980	3.61	4.46	3.85
1985	3.63	4.48	3.90
Western Seaboard			
1970	3.43	4.43	4.14
1975	3.42	4.43	4.16
1980	3.43	4.46	4.19
1985	3.46	4.48	4.23

3.2.4. LNG Costs — U.S.

The use of LNG as a sulfur-free fuel for power generation in a 10,000-MW power block is technically feasible. The combustion characteristics of vaporized LNG in an electric power plant are identical to those of domestic pipeline gas. The feasibility of using LNG as a sulfur-free fuel for power generation is based on its cost.

The 10,000-MW power blocks along the East and/or West Coasts of the United States to be served by LNG were assumed to operate at an annual load factor of 90%. The power blocks, therefore, produce 78.84 billion kWhr/yr of electric power. At an assumed heat rate of 9400 Btu/kWhr, this would result in 741,096 billion Btu/yr of fuel, or an average usage of 2030.4 billion Btu/day.

Four sources of LNG for the East Coast and two sources of LNG for the West Coast were studied. The East Coast sources — Venezuela, Algeria, Libya, and Nigeria — were selected on the basis of the availability of natural gas for export. The Middle Eastern sources of gas were considered beyond the economical distance for delivery to the U.S. The two West Coast sources of LNG were northwest British Columbia in Canada and Kenai pennisula in Alaska.

3.2.4.1. Liquefaction Plant Costs

3.2.4.1.1. Investment Costs

The costs of liquefaction facilities were estimated using existing cost information and projections of future trends. The estimated investment for the liquefaction plants was based on the relative severity of the refrigeration requirements — as measured by the latent and sensible heats of condensation of each source's gas — and the savings resulting from the magnitude of the operation. Cascade liquefaction systems with capacities of about 400 million SCF/day — the largest built to date — would cost about $125 million, or about $312.5/1000 CF liquefied per day. Single refrigerant systems (ARC cycle or similar) in the capacity range under study — several times larger than any built so far — would benefit from a considerable cost reduction due to the larger size and simpler cycle. The estimated investment for the liquefaction plants, sized to meet the needs of a 10,000-MW block, at each gas source was based on a relative severity of the refrigeration requirements — as measured by the latent and sensible heat needed to liquefy each gas — and the savings resulting from the magnitude of the operation. The unescalated (1969) costs for each of the six sources is shown in the following Table 3.2.-6.

Table 3.2.-6. LIQUEFACTION PLANT INVESTMENT

LNG Source	Plant Capacity, 10^6 SCF/day	Base Cost, $/$10^6$	Escalated Cost, $/$10^6$
Venezuela	2500	475	545
Algeria	2320	445	510
Libya	2080	395	455
Nigeria	2850	535	615
Canada	3000	545	625
Alaska	2860	535	615

The base cost includes the cost of the docking and port facilities and 1,200,000 bbl of LNG storage to allow rapid loading of the tankers.

The capacity of the liquefaction system was estimated using an energy flow balance (Table 3.2.-7.). (The energy flow balances for each system are presented in Section 7.2.3.2., LNG Cost Data.)

3.2.4.1.2. Operating Costs

The major operating costs for the liquefaction plant were estimated in four segments: fuel, labor and maintenance, depreciation, and fixed capital charges. The fuel required by the liquefaction plant has been estimated as 15% of the liquefaction plant daily flow. The labor and maintenance (including supplies) was estimated at 5% of plant investment and assumed to increase at the rate of 3% per year. Table 3.2.-8 details fuel requirements as well as the 1975 operating costs for a liquefaction plant sized to supply the fuel for a 10,000-MW power block. Depreciation was computed on a 20-year straight-line basis.

Table 3.2.-7. ENERGY BALANCE FOR LNG TRANSPORTATION SYSTEM

Annual ex-ship LNG deliver: 760 X 10^{12} Btu
One-way shipping distance: 1890 nautical miles

	Flow	10^{12} Btu per Calendar Year	10^9 SCF per Operating Day
1.	LNG Delivered by 8 Tanker(s) at 620,400 Barrels Each	760.0	1.89
2.	Tanker Fuel	34.1	0.08
3.	Net LNG Received by Tankers (1 + 2)	794.1	1.97
4.	LNG Gasified During Ship Loading at 1.5% of Net Lading	11.9	0.03
5.	LNG Pumped From Shore Storage Tanks (3 + 4)	806.0	2.00
6.	Boil-Off From Shore Storage Tanks (0.001/day) of 1,200,000	1.7	0.004
7.	Normal Liquefaction Plant Production Into Storage (5 + 6)	807.7	2.01
8.	Normal Liquefaction Plant Fuel Consumption at 15.0% of Production	121.2	0.30
9.	Normal Feed Gas Flow to Liquefaction Plant (7 + 8) − (4 + 6)	915.3	2.28
10.	Design Maximum Liquefaction Plant Production Into Storage (7 X 110%)	888.5	2.21
11.	Design Maximum Feed Gas Flow to Liquefaction Plant (9 X 110%)	1006.8	2.50
12.	Design Plant Production Capacity in 10^6 SCF	772.6 X 10^9 SCF/yr	2.21

(#10 X $\left(\frac{10^3}{\quad}\right) \left(\frac{CF}{Btu}\right)$)

Table 3.2.-8. LIQUEFACTION OPERATING COSTS

LNG Source	Fuel, 10^{12} Btu/yr	Operating Costs,* $\$10^6$
Venezuela	121.2	27.25
Algeria	127.2	25.5
Libya	127.45	22.75
Nigeria	132.0	30.75
Canada	121.5	31.25
Alaska	122.6	30.75

* Does not include cost of fuel for liquefaction.

Capital fixed charges were estimated at 20% overall. This is based on an assumed 50:50 debt/equity ratio. The interest rate on the debt was estimated at 10%. Equity was assumed to require a 30% rate of return with taxes.

3.2.4.2. Transportation Costs

3.2.4.2.1. Investment Costs

Transportation of LNG to the port of entry in the U.S. requires the use of specially built cryogenic tankers. Currently (late 1968) four such tankers are in service and six are under construction. These tankers range in size from 2000 bbl of LNG (in service) to 450,000 bbl (under construction). In addition, 650,000-bbl LNG tankers have been designed and negotiations are under way for their construction.

The number, size, and fuel requirements of the tankers required to deliver the desired quantity of LNG needed to fuel the 10,000-MW power block, to the East and West Coast were estimated. The capacity varied from 608,000 to 648,200 bbl of LNG. Table 3.2.-9 summarizes the overall requirements. The costs are based on cryogenic tankers powered with two gas turbines using vaporized LNG and operating at an average speed of 22 knots. Sufficient LNG storage capacity is included to meet the delivery and fuel requirements from each port of embarkation.

Table 3.2.-9. TRANSPORTATION INVESTMENT COSTS

LNG Source	Tanker Capacity, 1000 bbl	No. of Tankers	Total 1969 Investment	Total 1975 Investment
			$ million	$ million
Venezuela	648.2	8	288	330
Algeria	639.75	16	568	650
Libya	635.0	16	560	640
Nigeria	631.7	22	770	885
Canada	608.0	9	306	385*
Alaska	643.65	10	360	450*
Alaska (U.S. bottoms)	643.65	10	504	630*

* Includes strengthening for ice in northern waters.

For the Alaska-U.S. West Coast supply system both foreign and U.S. bottom (U.S.-built ships) costs were estimated. U.S. bottoms were assumed to cost about 40% more than foreign construction. The National Maritime Act requires all intercoastal U.S. shipping to be in U.S. bottoms with U.S. crews. For estimating purposes only, we have estimated both U.S. and foreign bottoms and crews. (Political considerations of the Jones Act will be discussed in Section 4.0.)

3.2.4.2.2. Operating Costs

The tanker operating costs consist of fuel, crew expenses, ship operating expenses, repairs, and port charges. The port charges consist of pilot charges, towing (tug boats), and port taxes. Table 3.2.-10 details the estimated port charges. Taxes are based on the New York Port Authority charges of $0.06/DWT*/trip (maximum annual charge $0.36/DWT/year). The tankers are rated at 100,000 DWT each.

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* Dead-weight ton.

Table 3.2.-10. PORT CHARGES
(Inbound or Outbound)

Pilot (U.S.)	$ 365
Towing (U.S.)	$1500
Foreign Port	$ 635
	$2500 X 2 = $5000/trip

The costs of the crew (45 men) subsistence, ship operating expenses, insurance, maintenance, etc., are detailed in Table 3.2.-11.

Table 3.2.-11. MONTHLY SHIP OPERATING COSTS
(Greek Crews)

	$
Wages, Overtime, and Benefits	13,800/ship
Subsistence	2,700
Deck, Engine, Stewards Dept.'s Operating Costs	9,000
Insurance	40,000
Claims and Repatriation	2,500
Repair and Maintenance Reserve	15,000
Misc., Travel, Communications, etc.	1,500
Management	8,500
Total Monthly Expense	93,000/ship
Annual Expense, ex-fuel	1,116,000/ship

The above are based on current costs. It is estimated that in 1975 the annual charge will be $1,160,000/ship. The cost of ships in service in northern waters — Canada and Alaska — must also include additional ice hazard insurance charges of 50%. The annual charge for these ships is $1,400,000/ship (1975 basis).

Those ships using U. S. crews will also cost more to operate. U. S. ships with U. S. crews (41 men) will cost about $2 million/ship/yr. The difference in crew sizes (45 Greek vs. 41 U. S.) is due to the absence of cadets on U. S. ships. The overall operating costs from each LNG source are shown in Table 3. 2. -12..

Table 3. 2. -12. ANNUAL TRANSPORTATION OPERATING COSTS

LNG Source	Tanker Operating Costs	Port Charges	Total Costs
	$ 10^6		
Venezuela	9. 28	1. 848	11. 128
Algeria	18. 56	2. 256	20. 816
Libya	18. 56	1. 728	20. 288
Nigeria	25. 52	2. 607	28. 127
Canada	10. 86	2. 349	13. 209
Alaska	12. 04	2. 110	14. 150
Alaska (U. S. bottoms)	20. 00	2. 110	22. 110

The ships are depreciated over 20 years on a straight-line basis. The overall capital charge is 10% on equity and debt. Current market conditions (December 1968) indicate an 8-3/4% rate on shipping for 10-15 year (maximum) financing. As a result of the long-term charter requirements and the stability of the overall program, it seems reasonable that 20-year financing will be available at an interest rate of 10%. Again, because risk is lower, equity capital should also be available at 10%.

3. 2. 4. 3. Storage Depot Costs

3. 2. 4. 3. 1. Investment Costs

Storage is required at both the LNG source and the East and/or West Coast port of entry. At the LNG source, a minimum level of storage is one full tanker cycle, or about 600, 000-700, 000 bbl of LNG. As a safety factor, the LNG storage at the source was set at 1, 200, 000 bbl.

The receiving terminal at the U. S. port of entry will require a similar amount of storage, about 600, 000 bbl minimum. In addition to the receiving terminal storage, sufficient LNG inventory must be maintained to guarantee fuel availability during periods of maximum power production. The level of fuel importation was set to meet the average (90 %) power production level of 216 million kWhr /day. The fuel required to meet the additional 24 million kWhr /day (57, 500 bbl of LNG /day) must be in inventory. The maximum storage requirement was set at that required to meet 125 days of continuous operation at 100 % electrical production load factor. The current estimated cost of LNG storage, in quantity, for large tanks was estimated at \$6 /bbl. The total installed cost, including piping, dikes, roads, and fences was estimated at \$7 /bbl. Table 3. 2. -13 details the fuel depot and receiving terminal storage requirements.

Table 3. 2. -13. STORAGE INVESTMENT

LNG Source	U. S. Fuel Storage Requirement, 1000 bbl	1969 Storage Investment Cost	1975 Storage Investment Cost
		$ 10^6	
Venezuela	7787. 5	54. 5	62. 3
Algeria	7887. 5	55. 2	63. 1
Libya	8037. 5	56. 3	64. 3
Nigeria	8075. 0	56. 6	64. 6
Canada	9025. 0	63. 2	72. 2
Alaska	8500. 0	59. 5	68. 0

The cost of the storage at the LNG source was included in the liquefaction plant cost. It was estimated at \$10 /bbl.

3. 2. 4. 3. 2. Operating Costs

The cost of operating the LNG storage depot cannot be separated from the cost of operating the receiving terminal /vaporization plant. The cost of operating the entire U. S. LNG facility is presented in Section 3. 2. 4. 4. 2.

3.2.4.4. U.S. LNG Facilities Costs

3.2.4.4.1. Investment Costs

In addition to the storage tanks, the U.S. facilities would include the port and docking areas, and the LNG vaporization system. We have estimated that the LNG-fueled power plant's share of the port facilities would cost $15 million.

The vaporization system, adequate to supply gaseous fuel to a 10,000-MW power block would cost between $3 and $5 million. The actual cost would depend on the source of the LNG and the capacity (measured in Btu's) of the equipment. The units have been assumed to be LNG-fired. Sufficient LNG has been imported. In addition to the vaporization units, fuel storage depot, docks and port facilities, and land (250 acres at $10,000/acre), some allocation for miscellaneous expenses has been included. Table 3.2.-14 details the estimated 1975 costs associated with receiving LNG from the various sources.

Table 3.2.-14. U.S. FACILITIES INVESTMENT

LNG Source	Vaporization Investment	Fuel Depot Investment	Dock and Port Facilities	Land and Misc Facilities	Total Investment
	$ 10^6				
Venezuela	4.0	62.3	15.0	6.7	88.0
Algeria	4.0	63.1	15.0	6.7	88.8
Libya	3.1	64.3	15.0	6.7	89.1
Nigeria	4.1	64.6	15.0	6.8	90.5
Canada	4.2	72.2	15.0	7.0	98.9
Alaska	4.4	68.0	15.0	7.0	94.4

3.2.4.4.2. Operating Costs

The cost of operating the U.S. facilities is detailed in Table 3.2.-15.

Table 3.2.-15. U.S. FACILITIES ANNUAL OPERATING COST

LNG Source	Labor and Overhead	Power	Maintenance	Taxes and Insurance	Total
	$ 1000				
Venezuela	870	871.2	880	1760	4381.2
Algeria	870	872.1	888	1776	4406.1
Libya	870	796.4	891	1782	4339.4
Nigeria	870	887.9	905	1810	4472.9
Canada	870	956.7	989	1978	4793.7
Alaska	870	918.4	944	1888	4620.4

The labor charge is based on 40 men including unloaders, pumpmen, firemen, control room operators, terminal engineers, and terminal managers. The general plant overhead was estimated at 100% of the labor cost. The power cost was estimated at 7 mills/kWhr for 77 kWhr/million CF/day vaporized plus an allowance for miscellaneous power (lighting, controls, etc.). The maintenance cost was estimated at 1% of facility investment. Local taxes and insurance were estimated at 2% of facility investment. Fuel costs were included in the cost of gas purchased at the source.

3.2.4.5. Refrigeration Recovery

The available refrigeration that could be recovered during the vaporization of LNG is approximately 280 hp/million CF of gas revaporized. The simplest recovery method, as shown in Figure 3.2.-4, is to use the vaporizing LNG to cool an intermediate fluid that flows countercurrently through a heat exchanger. The cooled fluid — methane or propane are likely candidates— is pumped up to high pressure, vaporized, and let down over an expander. While this is the simplest process, it is also the least efficient. If methane is used as the intermediate fluid and the LNG

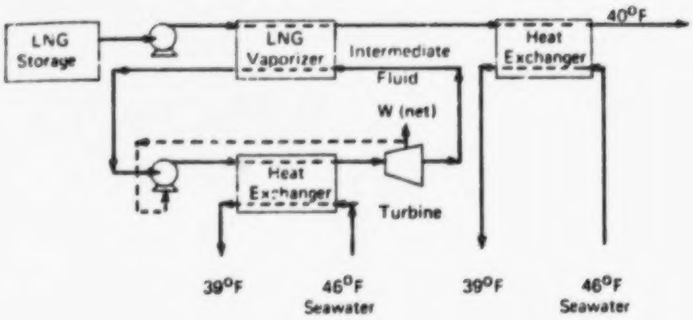
LNG Storage
LNG Vaporizer
Intermediate Fluid
Heat Exchanger
40°F
W (net)
Heat Exchanger
Turbine
39°F
46°F Seawater
39°F
46°F Seawater

Figure 3.2.-4. LNG VAPORIZATION — POWER GENERATION UTILIZING INTERMEDIATE FLUID POWER CYCLE[1]

discharge pressure is 14.7 psia, only 26% of the Carnot cycle work can be recovered.

The efficiency can be improved somewhat if the vaporized LNG is let down directly over the expander as shown in Figure 3.2.-5. In this case about 49% of the Carnot cycle work can be recovered.

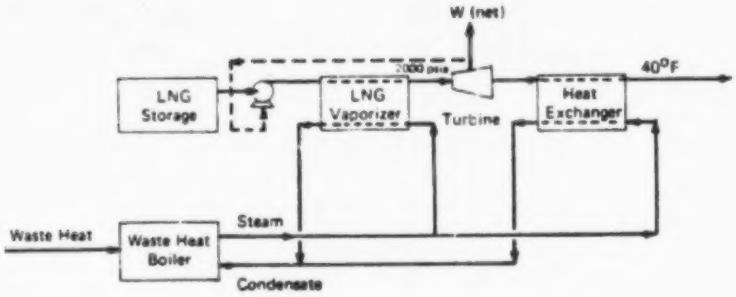
W (net)
2000 psia
40°F
LNG Storage
LNG Vaporizer
Turbine
Heat Exchanger
Steam
Waste Heat
Waste Heat Boiler
Condensate

Figure 3.2.-5. LNG VAPORIZATION — POWER GENERATION BY DIRECT EXPANSION OF VAPORIZED LNG UTILIZING AVAILABLE WASTE HEAT[1]

Since the energy recovery system would be operated in conjunction with an electric power plant, the recovered energy could be used to supplement the plant output. Table 3.2.-16 shows the added electric power that could be developed using either Cycle 1 or Cycle 2.

Table 3.2.-16. ELECTRIC POWER PRODUCTION INCREASE THROUGH ENERGY RECOVERY

	Cycle 1	Cycle 2
	%	%
Theoretical	5	5
Actual		
Vaporized LNG Discharge Pressure, psig		
14.7	1.3	2.5
140.0	0.6	0.9
300.0	0.3	0.7

Even though the potentially recoverable energy is only a small part of that used to liquefy the gas in a 10,000-MW installation, 2.5% added capacity is 250 MW.

The important question is whether or not refrigeration recovery by this method is economical. Various estimates[1,2] have placed the cost of recovery equipment at from $200 to $600/kW. It is to be expected that any energy recovery system will cost more than a straight gas-fueled power plant. Thus, to be economical, the energy savings must more than offset the added equipment cost. Using the same basis that was used in determining bus bar electric costs, that is, a 14%/yr fixed charge rate and a 90% load factor, it was determined that recovery equipment could only cost $280/kW more than conventional power generating equipment when fuel is valued at 50¢/million Btu.

In view of the uncertainties of equipment cost estimates, cold recovery for power generation was not considered in establishing bus bar electricity costs for the LNG-fueled plant.

There are many other attractive alternatives for cold recovery — alternatives which could significantly reduce the cost of imported LNG. For example, if the LNG refrigeration could be used in a liquid air plant, it is estimated that liquid air plant operating costs could be reduced by about 30%.

Power generating cycles could benefit directly from the use of LNG's cold. The conductivity of copper is 3.7 times greater at −260°F than at ambient temperatures. Cooling power plant equipment could greatly reduce copper requirements and, thereby, plant costs. The economics of the MHD power generating cycle could be improved if the necessary refrigeration for magnet cooling is obtained from the LNG fuel supply.

However, most of the applications pose formidable marketing problems. The quantity of refrigeration available is so great that its full use could probably never be realized in an industrial gas plant or chemical processing plant. Therefore, one of the most attractive uses is a combination plant in which the cold would be used for water desalination. The recovered cold from an LNG program sized to support 10,000 MW of electric power generation could desalt 150 million gal of water per day. While definitive cost estimates have not yet been prepared, it is felt that this might result in a credit of from 1¢ to 2¢/1000 CF to the imported LNG.

The potential for cold recovery to improve LNG importation economics merits much more extensive examination.

3.2.4.6. Total Gas Costs

The total gas costs were computed for a 20-year period starting in 1975 using the estimates presented in Sections 3.2.4.1. to 3.2.4.4. In each case the operating costs were escalated to reflect the basic changes in costs.

The delivered gas costs were computed using a computational procedure developed by IGT for the IBM·1800 computer. The program computes the total gas costs at the source, the total liquefaction plant costs, the ocean transportation costs, and the receiving terminal/fuel depot/vaporization costs.

The IGT-developed computation procedure determines the "levelized" cost of gas for gas delivered, stored, and vaporized for various assumed LNG sources and purchase prices.* These costs were developed assuming LNG delivery could start by 1975. Table 3.2.-17 presents the alternatives evaluated for both the Eastern and Western Seaboard plant locations. These gas costs are used in place of the domestic fuel costs presented in Section 3.2.3., Operating Costs, to determine electric generation costs using natural gas from imported LNG.

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* Levelized cost is a constant cost that produces the same present value as a steady stream of continuously increasing cost.

Table 3.2.-17. IMPORTED LNG COSTS

LNG Sources	Foreign Gas Cost, ¢/10^6 Btu	Vaporized Cost, ¢/10^6 Btu
Eastern Seaboard Costs		
Venezuela	5	40.0
	10	46.1
Algeria	6	48.1
	8	50.7
	10	53.3
Libya	6	45.5
	8	48.1
	10	50.7
Nigeria	6.6	59.5
	8	61.4
	10	64.1
Western Seaboard Costs		
Canada	6	46.3
	8	48.8
	10	51.3
	12	53.7
Alaska	10	52.3
	15	58.5
Alaska (U.S. Bottom)	15	63.4

3.2.4.7. Power Costs

The summation of fixed charges and operating costs presented in Sections 3.2.2., Investment Costs; 3.2.3., Operating Costs; and 3.2.4., LNG Costs, provides a comparison of total electrical generation (bus bar) costs for all four alternative fuel schemes —domestic natural gas, natural gas from imported LNG, oil, and coal. This comparison is presented in Table 3.2.-18. These costs consider the assumed investment and operating costs for desulfurizing the flue gases of oil and coal power plants.

Table 3.2.-18. BUS BAR COSTS FOR 10,000-MW ELECTRICAL GENERATING PLANTS
(Considers the Investment and Operating Costs for Desulfurizing Oil and Coal Flue Gases)

	Vaporized LNG Alternative Cost, ¢/10⁶ Btu								
Year	40.00	45.5	46.1	48.1	50.7	64.1	Gas	Oil	Coal
	mills/kWhr								
	Eastern Seaboard Costs								
1970	--	--	--	--	--	--	5.30	6.83	6.54
1975	6.09	6.59	6.65	6.84	7.07	8.30	5.54	7.14	6.92
1980	6.37	6.87	6.92	7.10	7.33	8.54	5.83	7.52	7.37
1985	6.70	7.19	7.23	7.42	7.64	8.83	6.17	7.95	7.90

	Vaporized LNG Alternative Costs, ¢/10⁶ Btu								
	46.3	48.8	51.3	52.3	58.5	63.4	Gas	Oil	Coal
	mills/million Btu								
	Western Seaboard Costs								
1970	--	--	--	--	--	--	4.94	6.65	6.72
1975	6.47	6.70	6.94	7.03	7.60	8.04	5.16	6.96	7.09
1980	6.72	6.94	7.17	7.27	7.82	8.26	5.43	7.34	7.53
1985	7.04	7.26	7.48	7.57	8.11	8.55	5.77	7.77	8.05

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The bus bar electricity costs for 1970 have been omitted as it is improbable that imported LNG for such a scheme could be available before 1975. However, cost comparisons for the other alternatives are presented for 1970. Table 3.2.-19 summarizes the break-even cost for imported LNG, that is the cost at which LNG can produce electricity at as low a cost as the lowest cost alternative.

Table 3.2.-19. BREAK-EVEN VAPORIZED LNG COSTS

Year	Vaporized Cost, ¢/10^6 Btu
Eastern Seaboard	
1975	49.0
1980	51.1
1985	53.6
Western Seaboard	
1975	51.6
1980	53.2
1985	54.7

As shown in Table 3.2.-18, electrical energy generated from LNG can be competitive with desulfurized coal and oil under many of the estimated LNG costs. These LNG costs cover both the high and low vaporized LNG cost alternatives and four of the intermediates presented in Table 3.2.-17. These final vaporized costs reflect several assumed foreign purchase costs. Many of the higher purchase cost alternatives are competitive. For instance, on the East Coast 10¢ Venezuelan gas, which has a 46.1¢/million Btu vaporized cost, would produce a competitive fuel. Likewise, 8¢ Libyan gas, which becomes 48.1¢ vaporized gas, would be competitive. On the West Coast 10¢ Canadian gas, which becomes 51.3¢/million Btu vaporized gas, would be competitive, and 10¢ Alaskan, which becomes 52.3¢/million Btu vaporized gas, would be competitive before 1980. Figures 3.2.-6 and 3.2.-7 show how various LNG supply schemes compete with oil and coal on the East and West Coasts. Inherent in the cost comparison of coal, oil, and LNG is a break-even cost of LNG. These costs are presented in Figures 3.2.-6 and 3.2.-7.

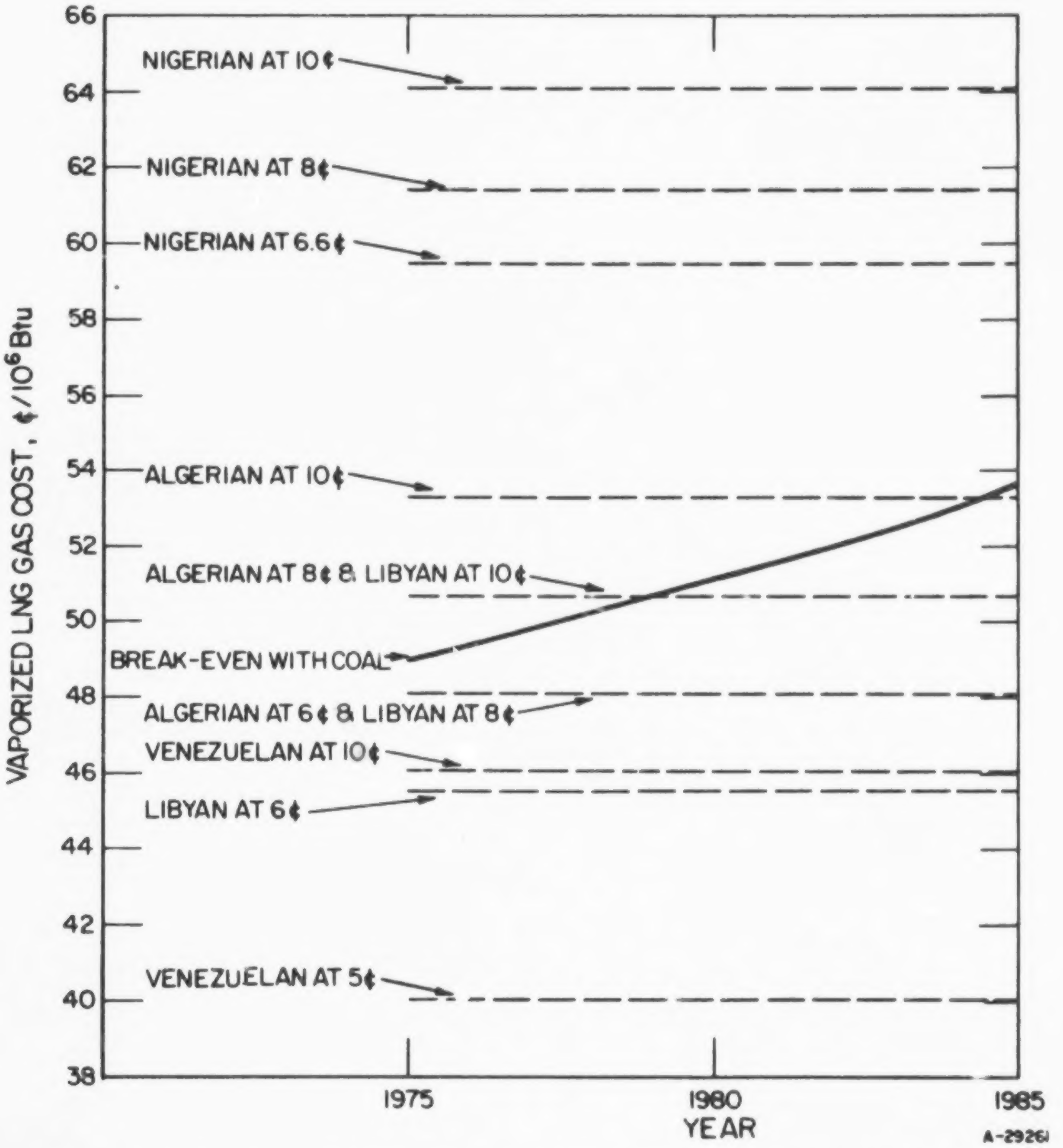

Figure 3.2.-6. EAST COAST LNG SUPPLY ALTERNATIVES

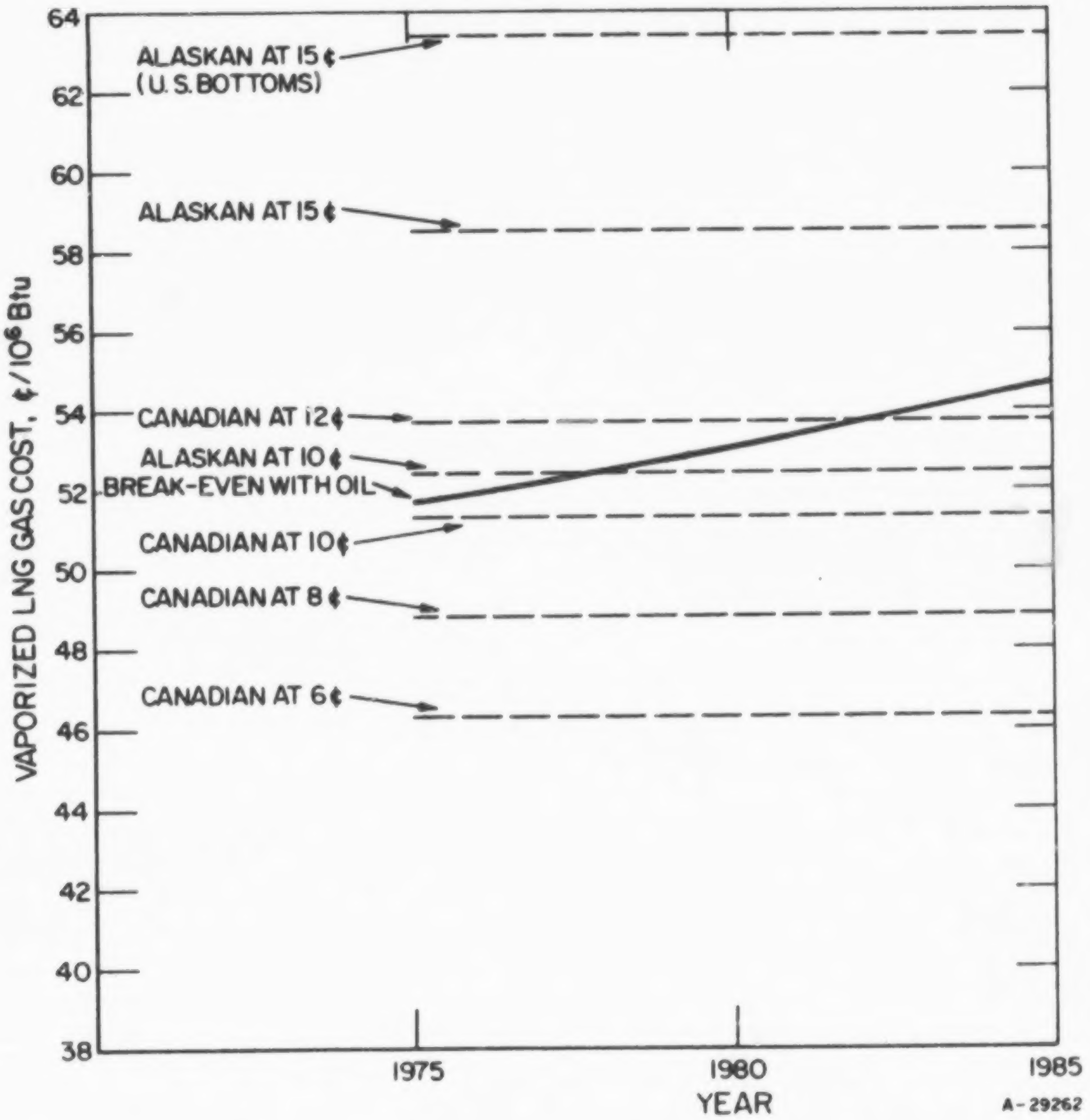

Figure 3.2.-7. WEST COAST LNG SUPPLY ALTERNATIVES

On the East Coast, with the exception of natural gas, coal will produce the lowest cost electricity; therefore, the break-even curve in Figure 3.2.-6 was made using the coal bus bar costs in Table 3.2.-18. On the West Coast oil will produce the lowest cost electricity; thus the oil bus bar costs were used to establish the break-even curve in Figure 3.2.-7. From these comparisons it is apparent that LNG-fueled electric plants can be highly competitive with oil and coal power plants.

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Section 4.0. Survey of Projection of Shifts in Prime Energy Source Utilization Patterns; Review of Possible Economic, Social, and Political Changes Resulting From Use of LNG as a Clean Fuel for Power Generation

4.1. Introduction

The pattern of energy utilization for power generation over the next 2 decades has been the subject of extensive studies that have resulted in many projections. However, none of these projections has seriously considered the impact of imported LNG as a possible energy source. The recently released Federal Power Commission Report on Air Pollution and Regulated Electric Power and Natural Gas Industries pointed out that "imports of liquid natural gas may become more important in the future making the United States less dependent on North American reserves."[5] However, it made no quantitative assessment as to the possible inroads of LNG into the power plant market. Indeed, it is difficult to quantify such inroads in view of the currently changing status of such important market factors as pollution control regulations, import regulations, competition for world gas supplies, and even the domestic gas supply problem.

There is no doubt that there will be a greatly increased demand for energy for electric power generation through the rest of the 20th Century. The convenient rule of thumb of doubling of power output every 10 years may lead to an erroneously high estimate of electric power generation by the year 2000, but, nevertheless, growth will be tremendous. There are no indications that there will be radical changes in the energy sources used for power generation, except for the well-publicized, and readily accepted, advent of nuclear energy. Coal, natural gas, and oil, along with hydropower, will continue to supply a major portion of the energy for power generation. And, although, by the decade of the 1980's, gas, the fossil fuels, and hydropower may well have surrendered a significant share of the power generating market to nuclear energy, the total quantity of gas consumed in this market segment will, in fact, be larger than the current consumption of 2.6 trillion CF. One recent estimate by Humble Oil and

Refining Co.[3] assumed a growth rate of 2.8%/yr for gas used by electric utilities, indicating that by 1985 over 4.0 trillion CF of natural gas would be used by electric utilities. The current rate of growth of this market, responding at least in part to the effects of air pollution regulation, certainly indicates that this estimate was not too high. While the growth rates will vary in different regions, growth will be much more rapid in the gas-competitive areas of the Southwest and South, and somewhat slower in the high-fuel-cost areas where nuclear energy can more readily encroach upon the market. Nevertheless, the total effect will be a need for increased quantities of natural gas for power generation.

The source of this gas, whether the indigenous fields of the United States, imported gas from adjacent North American countries, or sources around the world as LNG, will be affected not only by the economics of the market place and interfuel competition but also by the many regulations imposed on the transport and use of energy. LNG, in particular, is possibly subject to a broad range of regulatory requirements, many of which have not, as yet, been clarified. The Federal Power Commission, acting as the agent for the President of the United States, must approve import of natural gas; the Oil Import Administration, even though it has expressed the opinion currently that LNG is exempted from its regulation, may, at some future date, be placed in some position of control. There are continuing efforts to place a duty on imported LNG, a move that would involve the Customs Office. The Federal Maritime Commission is concerned with transport of LNG, in particular with the administration of the National Maritime Act, the so-called Jones Act, which greatly affects interstate transport of LNG (a significant concern in the possibility of transporting Alaskan natural gas to other states); the United States Coast Guard is concerned with the safety of transport and handling to ensure the safety of other vessels in the United States' harbors.

Because of the newness of LNG technology and the unfamiliarity of many of the regulatory groups withit, no body of expertise has yet been developed which can provide the base for defining the regulations in light of the importation of LNG. Nevertheless, it is possible to make some qualitative estimates of the impact of liquefied natural gas on the various regulatory agencies and the competing fuels for power generation.

4.2. Nuclear Energy

Although liquefied natural gas is proposed as a low-pollution alternative to high-sulfur fuels such as coal and residual oil, no doubt its importation would have some impact on the growth of nuclear energy for power generation. Gas-fueled power plants enjoy a significant first-cost advantage when compared to nuclear plants. With the present high interest rates, low first cost is a very great advantage for base-loaded electric utility plant operation, and, as plant factors drop, the lower cost plant is further advantaged. But nuclear energy will certainly enjoy a dynamic and rapidly growing future as a source of energy.

Nuclear energy for power generation is the beneficiary of an extensive government-funded research effort. Although the present converter reactor technology extracts the latent energy of the uranium atom in a relatively inefficient and, as yet, expensive process, it is, nevertheless, an almost unavoidable step along the path to a breeder reactor economy with its promise of almost unlimited, very low-cost electric energy. The massive research effort of the Atomic Energy Commission and industry can be expected to resolve the problems breeder reactors still face and eventually lead to a nuclear energy based electric generating industry.

Nevertheless, the nuclear power industry is faced with several problems, some of them of a minor and temporary nature, others longer term and more enduring. Presently, nuclear plant construction is slowed by the careful evaluation of all safety features. This is felt to be temporary. Once the regulatory process is smoothed out and greater standardization achieved, licensing and construction delays will probably be reduced. Part of this gain may be offset because siting problems, discussed in more detail later, are actually increasing rather than decreasing. But in any event, the administrative problems of regulatory approval are likely to be reduced.

It is axiomatic that nuclear power plants must be big to be economical. But this advantage in itself breeds other problems, particularly with plant siting and distribution of the energy from a large central source. Electric power cooperatives, for the most part so small that they cannot support the mammoth 1000 or 2000-MW nuclear plants, are seeking a way to share in the economies of scale of these size plants. They have sought legislation that would allow them to purchase power at what they consider to be a reasonable cost for use in their distribution systems.

Of these problems, perhaps the greatest problems at the present time have to do with plant siting, and not the least of these are the problems engendered by thermal pollution, i. e., discharge of waste heat to adjacent bodies of water. Since nuclear power plants lack the capability for using superheat cycles, they are inherently less efficient than fossil fuel plants, and, since none of the waste heat is rejected up the stack, thermal pollution may be as much as 60% higher than for fossil fuel plants.[7] Of the two alternatives — use of cooling towers and recirculated water or using a once-through cooling system — the former increases plant cost at some further economic penalty and the latter is said, by conservationists to create ecological problems. There appears to be little technical hope on the immediate horizon to offer relief from the thermal pollution problem. Nuclear power plants will have to be located at remote sites where they will have a minimum impact on the ecological balance or pay the penalty for cooling towers.

Air pollution in the commonly accepted context of sulfur or particulate matter pollution is not, of course, a problem with nuclear power plants. Neither do they ordinarily emit into the atmosphere any significant amount of radioactive material or nonradioactive gaseous or particulate matter. When a nuclear power plant is in operation some radioactive materials are formed — mixtures of gases and solids, such as iodine, krypton, xenon, and plutonium. However, these materials are almost entirely contained within the fuel, less than 0.001% escaping. The handling of fuel elements during reloading and shipment to the fuel reprocessing plant is carefully designed to eliminate any possible escape of radioactive products.

When the fuel is reprocessed, radioactive fission products are released from the fuel as a gas. The chemically reactive gases, such as iodine, can be collected on charcoal or in scrubbing solution, but nonreactive gases such as krypton and xenon are much more difficult to control. Xenon, possessing as it does a relatively short half-life, can be neutralized by radioactive decay. On the other hand, krypton 85 is discharged to the atmosphere. It presently poses no problem, but at some future time, as nuclear power generation increases, it may become a matter of concern. It has been estimated that, on the basis of projected nuclear power development by the year 2000, body exposure to krypton 85 would be approximately 0.01 of the limit recommended by the International Commission on Radiological Protection.[5]

Even though nuclear power plants are not faced with the air pollution problems of fossil fuel plants, they have some peculiar problems of their own which have great impact upon their location near urban centers. The safety of nuclear power reactors is rigidly controlled. An overwhelming body of information exists that indicates that nuclear power reactors are technically safe from any nuclear explosion. Nevertheless, this does not eliminate the concern felt by many people over the location of such a tremendous amount of energy close to population centers. However, the likelihood of "uncontrolled release " of this energy is so remote as to make it almost incredible. The impact of such a possibility is much more likely to be felt through its effect on the increased cost of reactor safeguards and delays in plant licensing. Research costs , improvements of materials and equipment, and continued testing to ensure their proper operation are also cost factors that must be equated against the problem of ensured complete safety.

The relative advantages and disadvantages of locating nuclear electric power plants close to the population centers has been debated at great length. Recently, a Citizens Advisory Committee advanced the argument that nuclear plants should be located close to the population centers to "eliminate the need for long-distance transmission of power from the generating center to the center of consumption."[2] The recently released Report to the President's Council on Recreation and Natural Beauty[10] appeared to refute this contention, in part, by pointing out that, because of their very large size, nuclear power plants will still require extensive transmission grids to interconnect the plants with other large plants and to disperse the energy over a wide market area. The report also pointed out the need for a continuing safety research program like the one being conducted by the Atomic Energy Commission to ensure the highest degree of safety for all nuclear power plants. This would tend to increase the likelihood of the ultimate acceptability of nuclear electric power plants in population centers. The report concludes its consideration of nuclear power plant siting with the comment:

> "Until there is more experience in the design, construction, and operation of nuclear plants, the Atomic Energy Commission plans to maintain a conservative approach evaluating plant safety and establishing a balance between the compensating engineering safety features and population density." [10]

It would appear that the Atomic Energy Commission will follow this path for the near-term; therefore, nuclear electric power plants may suffer some cost disadvantage longer if they must be located in areas more remote from population centers than fossil-fueled plants.

Nuclear fuel supply adequacy also continues to be a subject of protracted debate. While there is little doubt that technology will eventually overcome the immense problems of controlled fusion and that we will eventually have "unlimited energy," in the near term, there are serious limitations on nuclear fuel availability. Meeting the demand by developing high-cost fuel supplies can relieve a possible shortage, but only with a concomitant penalty against nuclear fuel economics.

The most recent assessment of the effect of nuclear fuel availability was recently presented by Rafford L. Faulkner, Director of AEC's Division of Raw Materials.

The starting point for Mr. Faulkner's forecast is AEC's December 1967 Forecast of Growth of Nuclear Power.[1] This report predicted that, by 1980, the installed nuclear generating capacity in the U. S. would be between 120,000 and 170,000 MWe. At the median point in this range, the cumulative uranium requirement from 1966 through 1980 was estimated at nearly 250,000 short tons of U_3O_8. The annual requirement in 1980 would be about 38,000 tons.

On the basis of this forecast, requirements in the 1975-1980 period would be met mostly by uranium ore reserves yet to be discovered. The AEC is optimistic that current high exploration rates will yield these necessary additional reserves. It also expects that countries having supplies in excess of needs, such as Canada and South Africa, will seek to market them in the United States.

The year of maximum uranium production in the United States was 1960-1961 when almost 18,000 tons of U_3O_8 were produced. The production in 1968 amounted to about 13,000 tons. Current production capacity is near 14,000 tons/yr. The AEC assumes that this could be increased to 20,000 tons/yr over a period of several years.

For the rest of the non-Communist world, nuclear fuel requirements through 1980 roughly equal those of the U.S. Its production, combined with U.S. production, is currently about 23,000 tons of U_3O_8 per year. Expansion to 40,000 tons by the mid-1970's is considered possible. However, the annual non-Communist world requirement during the 1975-1980 period will increase to about 75,000 tons/yr. Meeting this demand at costs acceptable to consumers will be a major undertaking. Mr. Faulkner acknowledges this, but says, "it seems likely to be within the capability of the industry."[4]

After 1980, uranium production requirements are expected to increase even further. The amount of increase depends not only on the more evolutionary economic trends but also on the progress of breeder reactor development and commercialization. Figures 4.2.-1 and 4.2.-2 illustrate the range of possibilities visualized by the AEC. Cumulative U_3O_8 requirements will increase from 250,000 tons in 1980 to about 1 million tons in 2000. This latter figure may vary + 20% depending on the breeding gain characteristics of reactors introduced into the power grid after 1980. The upper curve in Figure 4.2.-1 reflects the assumption that initial, oxide-fueled, low-performance breeder reactors with doubling times of 12-13 years will not materially improve during the 1980-2000 period. The lower curve is based on the assumption that higher performance and advanced breeder types ultimately achieving doubling times of 6-7 years will be introduced during this period. The nation's total installed nuclear-generating capacity in the year 2000 is assumed to be 735,000 MW, or approximately 50% of the total electric generating capacity.[1]

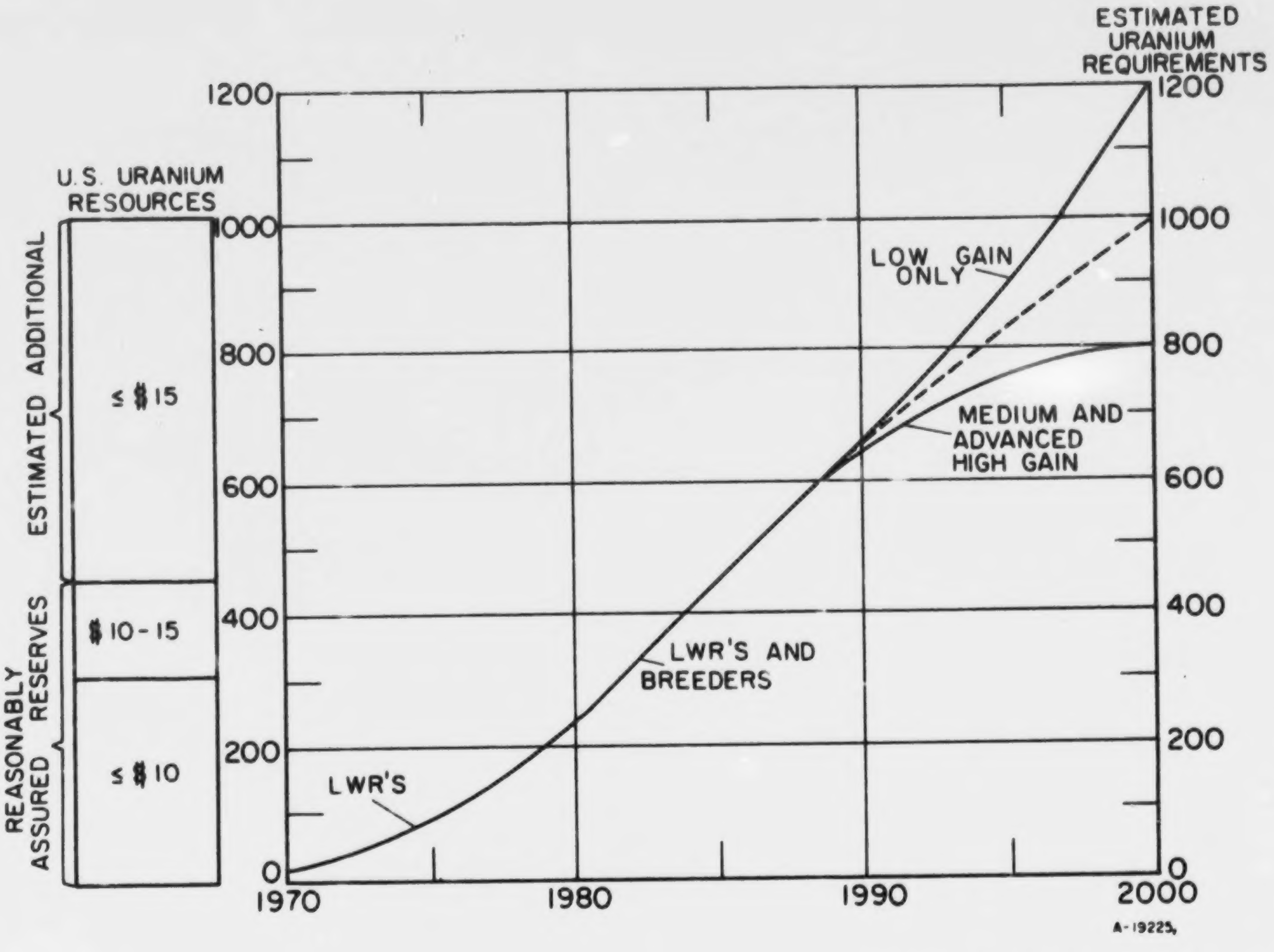

Figure 4.2.-1. U.S. CUMULATIVE URANIUM REQUIREMENTS PROJECTED THROUGH THE YEAR 2000[4]
(Short tons of U_3O_8)

The bar graph at the left side of Figure 4.2.-1 indicates that, if uranium prices of $15/lb (roughly double the current price) can be commercially accepted, the total reasonably ensured and estimated reserves in the U.S. will equal the cumulative needs by 2000 — referring to the median requirements estimate of 1 million tons. Of this amount, two-thirds is expected to be available at $10 or less per pound, and approximately one-half of the total remains to be discovered. It is recognized, of course, that if no more than 1 million tons of reserves were available,

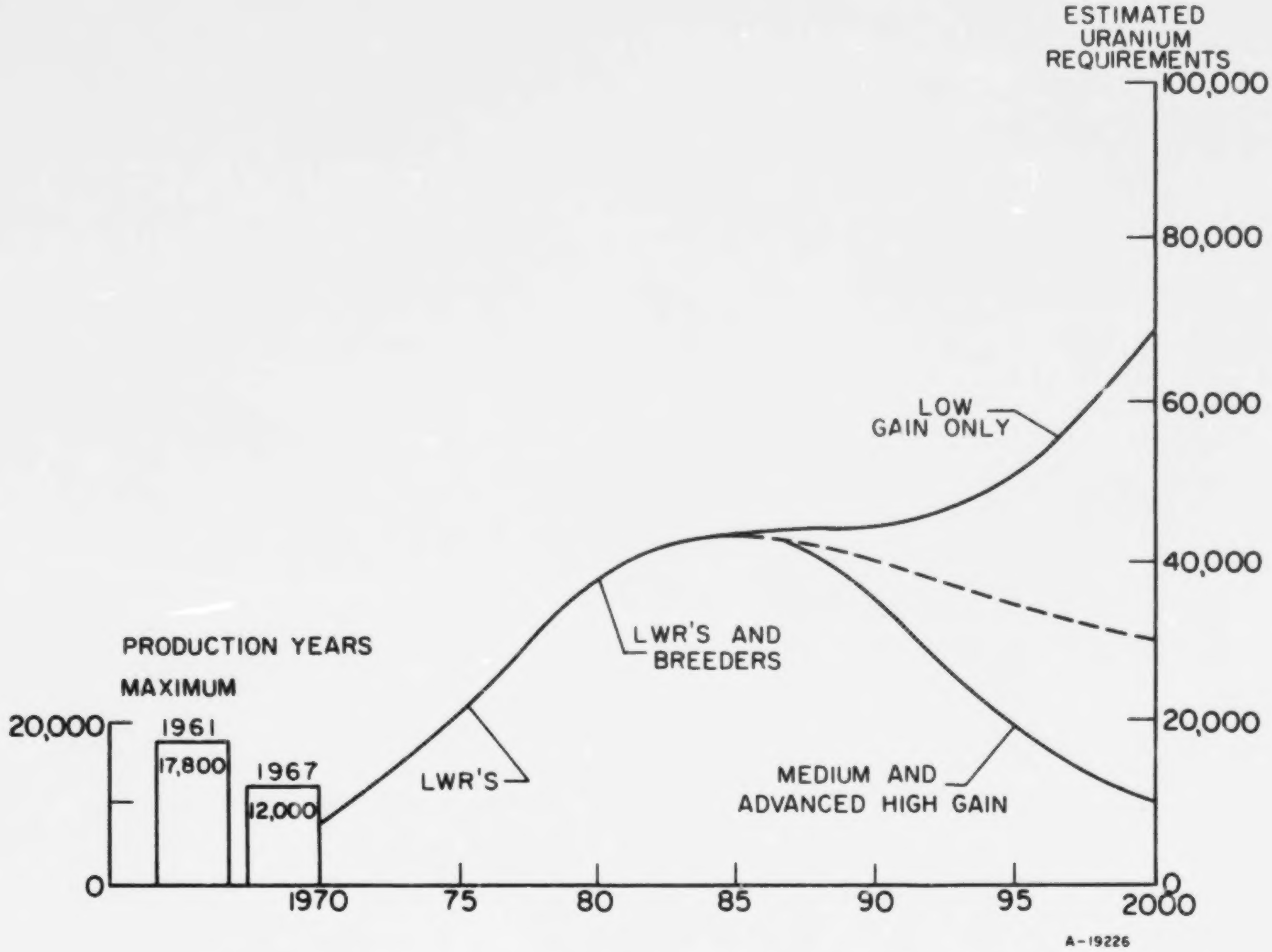

Figure 4.2.-2. U. S. ANNUAL URANIUM REQUIREMENTS PROJECTED THROUGH THE YEAR 2000[4]
(Short Tons of U_3O_8)

declining production rates in the later years would not support the projected median requirements; therefore, a substantial additional supply would be needed.

The estimated annual U. S. requirements, corresponding to the cumulative requirements of Figure 4. 2. -1, are shown on Figure 4. 2. -2. A careful comparison of Figures 4. 2. -1 and 4. 2. -2 shows two characteristics of the nuclear energy industry. First, there is an extremely wide range of uncertainty in requirements. The "pessimistic" (upper) curve of Figure 4.2.-1 shows requirements increasing at an ever-increasing rate, while the "optimistic" (lower) curve indicates that there will be virtually no need for any additional reserves after 2000. Thus, the ± 20% range in Figure 4.2.-1 for the year 2000 becomes a much wider range (in Figure 4.2.-2) in terms of required annual production. The reason for this is that, when a sufficient number of high-gain breeders come into service, not only the U^{235} component constituting 0.7% of natural uranium but essentially all of the natural uranium becomes a fuel. This means that spent fuel from nuclear reactors, tails from U^{235} separation plants depleted in that component, and all naturally occurring uranium (unenriched in the U^{235} component) become available as nuclear fuels. This addition to resources greatly diminishes or, for a period of years, totally eliminates the necessity for mining additional uranium ore. On the other hand, a nuclear economy based solely on low-gain breeders and burners must not only find new fuel for the new reactors being installed (presumably at an exponentially increasing rate) but it must also mine ores for refueling the ever increasing number of existing reactors. Therefore, the requirements estimate of 30,000 tons of new U_3O_8 for the year 2000 (median estimate from Figure 4.2.-2) probably has little validity, the actual figure depending on factors that cannot be properly assessed at this time.

In addition to the uncertainties involved, the other characteristic made evident by these projections is that there is, in fact, some critical time by which breeders should carry a heavy share of the nuclear generation burden if fission power technologies are to survive as a significant factor in the total energy generation picture. This is not to endorse the accuracy of any particular estimate of when that critical time will be, but merely to note the rapidly escalating difference between requirements and resources if at some point a breeder economy does not develop along with an expanding burner economy. Even if the usable supply of nuclear fuels were to be doubled in the year 2000, relative to the quantity assumed in Figures 4.2.-1 and 4.2.-2, the steep slopes of the upper curves in each figure make it evident that this would not delay the critical period by more than a decade or so. Thus, without a breeder economy, only moderate improvements can be made by developments such as finding ways to use more costly uranium ores economically, creating a thorium-based reactor technology, or discovering ore reserves far greater than those currently considered probable.

The requirement-supply situation for non-Communist countries as a whole is quite similar to that in the U. S. Cumulative requirements to the year 2000 may be 3 million tons (± 50 %), which are about equal to estimated reserves costing up to $15/lb, of which approximately 1.6 million lb should cost not more than $10/lb.

The importance of nuclear breeder reactor development appears to be well recognized by both Government and industry, and we see no great controversy existing as to the urgency of essentially completing this development in the 1980's. There has not, however, been much public discussion of the consequences of failure to achieve a timely breeder development, and such a possibility is a real one in view of the magnitude and complexity of problems to be solved. While it appears likely that all necessary effort will be applied to the timely development of practical breeders, the early development of the LNG importation industry would provide a most welcome form of insurance at a critical time in the nuclear power industry's development.

In the attempt to evaluate the shorter range (<20 years) impact of nuclear energy, both the improvement of nuclear conversion efficiency and the timing and extent of resources development assume considerable importance. There are also numerous uncertainties in predicting the course of ore recovery and refining developments. Judgments in this area are necessarily based largely on interpretations of current and past experience.

In view of past successes in finding the needed amounts of nuclear fuels, it would be easy to take a complacent attitude about future prospects, but it will not be easy to produce an average of 30,000 tons/yr of U_3O_8 over the next 30 years. Some of the factors that indicate a need for caution in assessing the future are —

a. Surface occurrences of uranium-bearing ores led to the discovery of most of the important deposits now known, but surface showings appear likely to become decreasingly important in the future.

b. As average drilling depth increases, a decreasing frequency of sampling of the target formations results from a given number of feet drilled; thus, the thickness and grade of the ore deposits must be greater to warrant development.

c. In the past 18 months, most of the drilling has taken place and much of the new ore has been found in and around districts of previous production. These districts contain about 90% of our reserves and have supplied nearly 70% of total production.

Several other factors support the idea that the needed reserves can be found. Many of the strong, well-financed companies are still making geological studies, but have yet to execute sizable drilling campaigns. Also, many of the potentially favorable areas which have not had significant production have not been sufficiently explored that they can be discounted.

The selection of which plant types are to be ordered for start-up in the 1975-1980 period must be among the more critical problems currently facing electric utility management. Large uncertainties exist in both the fuel supply and the fuel utilization ends of the business. The importance of making the right decisions could account for their postponing new commitments as long as possible to take advantage of the most up-to-date information being developed in all areas of technology, economics, and politics.

4.3. Coal

The second dominant influence in the energy market for electric power generation is coal. At the present time, coal provides approximately 52%* of the raw energy used for electric power generation, and almost one-quarter of the total energy required in the United States for all purposes. Coal is the most abundant of the fossil fuel reserves, with a total of 1576.2 billion tons classified as existing proved reserves. This figure must be drastically discounted, of course, for recovery efficiency. It is commonly considered that a 50% recovery factor is reasonable.

While these reserves are vast, present and anticipated pollution control regulations and the economics of transportation are likely to limit their utilization for power generation. For example, as shown in Table 4.3.-1, only 13.7% of the total reserves, or about 216 billion tons, could be categorized as low-sulfur bituminous coal, the "ideal" coal for power generation. While 65% of the coal reserves are classified as being low sulfur — that is, containing less than 1% sulfur — by far the largest proportion, the vast subbituminous and lignite reserves of the Western States, are remote from the large market areas. These reserves could be, and likely will be, utilized through the use of mine mouth plants with high-voltage electric transmission grids, but all of these factors will tend to increase the competitive cost of coal utilization for power generation.

But, even if low-sulfur bituminous coal replaced the higher sulfur content coal that has been used for steam raising, several other technological changes occur. It has been pointed out that there is a strong inverse relationship between the sulfur content of coal and its ash-softening or fusion temperature. As the sulfur content declines, fusion temperature increases. Therefore, when a lower sulfur fuel is used at the expense of a higher fusion temperature, heat exchanger fouling, and slag-tapping problems are intensified.

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* Derived from FPC data.

Table 4.3.-1. SULFUR COAL CONTENT OF U.S. COAL RESERVES[5]

	Low Sulfur (1% or less)		Medium Sulfur (1.1-3.0%)		High Sulfur (over 3.0%)		Total	
	Billion Tons	% of Total	Billion Tons	% of Total	Billion Tons	% of Total	Billion Tons	% of Total
Bituminous	215.7	13.7	194.9	12.4	314.2	19.9	724.7	46.0
Subbituminous	387.2	24.6	1.5	0.1	-	-	388.7	24.7
Lignite	406.0	25.8	41.6	2.6	-	-	447.6	28.4
Anthracite	14.7	0.9	0.4	*	-	-	15.2	0.9
Total	1023.6	65.0	238.4	15.1	314.2	19.9	1576.2	100.0

* Less than 0.05%
Figures may not add to totals because of rounding.

With the higher degree of ash carry-over, fly-ash precipitation equipment and ash-handling equipment would be increased. Also, there is some indication that the efficiency of existing electrostatic precipitators is adversely affected by changes in the physical characteristics of the fly-ash produced.

The total ash content is also often increased when the lower sulfur coals are substituted. This means that the fly-ash precipitators and other ash-handling equipment would be undersized in a plant that was converted to low-sulfur coal and would need to be increased in capacity at some further cost.

Other problems, not of as great significance, also occur: 1) Grindability characteristics often change, meaning that additional grinding mills may be required, and 2) Changes in coal volatility may necessitate changes in burner design or installation.

The coal-fired plants are probably the least culpable of all of the steam-electric plants insofar as thermal pollution is concerned. Plant efficiency is quite good, with heat rates in the range of 8500-8700 Btu/kWhr not uncommon in new plants. However, even with a 40% efficient plant — a heat rate of 8533 Btu/kWhr — 3800 Btu/kWhr would still be rejected to the cooling water. Coal-fired power plant heat rates seem to be leveling off, having reached the point of diminishing return on the use of reheat cycles. The single reheat 1050°/1050°cycle is most commonly used now. It appears difficult to justify the complexity of additional stages of reheat or the higher cost of the special materials that would be required for increasing reheat temperatures. However, extensive research effort is under way to improve plant efficiency to reduce both fuel costs and thermal pollution.

Techniques such as magnetohydrodynamics, electrogasdynamics, thermionic topping cycles, and numerous other direct energy conversion devices have been suggested as a means of raising coal-fired as well as other fossil-fueled electric power plant efficiencies to the range of 60%. The Office of Coal Research of the U. S. Department of the Interior is funding four separate programs that are intended to improve the efficiency of electric power generation. The combined interest of the coal industry, the coal miners' unions, the equipment manufacturers, railroads, and other groups vitally concerned with the advancement of the coal industry, which originally led to the formation of the Office of Coal Research, can be expected to sustain the research effort on finding methods to enhance the suitability of coal for power generation.

Coal-based power plants are also faced with the siting problems common to all of the steam-electric power plants. They do produce some thermal pollution and, therefore, have problems with water availability and the effect of the rejected heat on the environment. Land availability is a particularly significant problem to coal-based plants because the need for extensive and highly visible coal storage piles is a unique problem. Special provisions must be made to avoid liquid pollution runoff from coal storage areas and to eliminate the possibility of spontaneous combustion with its concomitant, uncontrollable air pollution.

But air pollution from both sulfur and particulate matter is the prime problem facing the coal industry. Recognizing this several years ago, the industry mounted a campaign to develop the means for combating this problem. Technically, the problem of fly-ash control is not difficult. Equipment such as electrostatic precipitators, or cyclones, can effectively reduce fly-ash emission. However, all of these impose some cost burden on the power plant. The industry has attempted to improve the economics of fly-ash control by developing markets for the reclaimed material. Fly ash does have a number of uses such as concrete aggregate. The electric utility and coal industry have combined to form an association[9] to promote the marketing of fly ash.

On the other hand, sulfur poses a much more formidable control problem. Since it is burned in the boiler to a gas, mechanical means are not effective in removing the sulfur after combustion.

Sulfur may be controlled by desulfurizing before combustion or by removal of SO_2 and SO_3 by chemical means from the stack gases after combustion. In the former methods, pyritic sulfur is present as separate particles and, in theory at least, could be separated mechanically from the coal in which it is contained. However, to do this requires fine grinding to break the coal matrix down to a size from which the sulfur particles can be separated. Once the coal is finely ground, separation can be accomplished by wet cleaning techniques, such as the use of washing jigs, dense media separation, concentrating tables, and flotation techniques. It has been estimated that conventional cleaning costs in a large coal preparation plant would be approximately 50¢/ton. To this must be added the cost of the necessary finer grinding.

In addition to techniques for applying the already available cleaning methods, research is under way on several other mechanical cleaning processes including electrostatic cleaning, magnetic cleaning, froth-flotation, concentrating tables, and air classification.

It has also been suggested that chemical methods might be used for the reduction of pyritic sulfur. The two methods that have received some degree of attention are chemical oxidation and bacterial oxidation. These involve conversion of the pyritic sulfur to a water soluble sulfate.

Thermal processes for sulfur reduction have also been suggested. Low-temperature carbonization has been applied for many years to convert a portion of coal to tar and gaseous products. Formerly, the tar was sold for its chemical value and the gas used for heating. Raw coal can also be solvent-refined by a solvation process. In this process a large portion of the organic sulfur is hydrogenated to hydrogen sulfide, which can be removed as a gas. The pyritic sulfur is partially converted to organic sulfur and the remaining pyritic sulfur is separated with the other mineral matter, the ash and the coal. Experimental results have indicated that as much as 70% of the total sulfur can be removed by this process.

While the mechanical and chemical methods of precombustion sulfur removal are under continual investigation, the Federal Power Commission has concluded in its analysis of fuels for power generation that these methods will, at best, result in only partial removal of the pyritic sulfur. The FPC states that "in view of recent forecasts of potential cost of removing sulfur from the flue gases or during combustion, elaborate cleaning or treating methods prior to combustion have little economic potential in the electric power industry except in the case of some established plants where addition of stack devices may not be feasible."[5]

Because of the difficulties of precombustion sulfur control, the principal emphasis has been placed on control of sulfur oxides. In general, these have involved the removal of the sulfur compounds formed during combustion before they are discharged into the atmosphere or the use of high stacks to dilute and disperse the pollutant. Tall stacks are effective by present standards, but are felt to be only a short-term solution.

In addition to the costs imposed by sulfur control, it is to be anticipated that the increasing competition for the low-sulfur coals, which will be allowable in the existing coal-fired power plants, will cause an upward pressure on prices. For example, in its quest for a low-sulfur coal approaching the specifications of a low-sulfur coking coal, one large Eastern utility was forced to pay $1.90/ton above the regular price of coal in the past.

Because the large reserves of low-sulfur coal are located in remote areas, it is anticipated that transportation costs will become an increasingly important segment of coal-based power costs. The technological and economic improvements in coal transportation costs brought about by the advent of the unit train and coal pipelines will be negated to a large degree by the much longer transportation distances involved. While extra-high-voltage transmission of power from mine mouth plants can be expected to eliminate some of the coal transportation costs, these costs are replaced in part, by the cost of transmitting the electricity.

Mining costs will certainly be under pressure from union demands for higher wages and general inflationary effects. The recently concluded United Mine Workers contract will result in an increase in coal prices estimated to be approximately 30¢-40¢/ton over the next 3 years. Since coal mining still has a very high labor input, wage increases very directly affect mining costs. Improved efficiency of mining, a higher degree of utilization of mining equipment, and new mining techniques can be expected to help hold the line on coal prices.

4.4. Oil

Oil provides the smallest share of the three fossil fuel sources of energy for electric power generation. Currently it supplies approximately 8%* of the raw energy required. It has been estimated that by 1985 this share of the market will decline to approximately 3%. Historically, the oil used for power generation has had a high sulfur content. Increasingly stringent air pollution control requirements have forced the substitution of low-sulfur oils at some penalty in price.

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* Derived from FPC data.

Low-sulfur oils for power generation, or for any other purpose for that matter, may be obtained in three ways: residual oil can be produced by refining low-sulfur crude oil, the oil can be desulfurized, or low- and high-sulfur oils can be blended. Power plants also have the option of scrubbing the SO_2 produced during combustion from the stack gases.

While it is not expected that natural low-sulfur residual fuel oil will be a significant factor in reducing the nationwide problem of sulfur dioxide pollution, it may make some marked contributions in certain regions. Principal sources of residual fuel oil for the United States are refineries in Venezuela and the West Indies, which process Venezuelan crude.

Although the importation of crude oil and petroleum products from countries other than Canada and Mexico is controlled by the Oil Import Administration, the importation of residual fuel to the East Coast is not restricted at the present time. The Oil Import Administration has also set up regulations for West Coast producers which will permit each eligible refiner to receive a crude oil import allocation equal to the amount, in barrels, of low-sulfur fuel oil supplies.

Techniques are available to desulfurize oil. Hydrogen treating methods, such as the Isomax, Texfining, H-Oil, and others, combined with blending of high-sulfur oils, are finding wide acceptance. All of these processes incur significant additional costs. It has been estimated that residual fuel oil can be desulfurized to 1% or less sulfur
early to predict future costs on the basis of technological breakthroughs, it has been estimated that process improvements and economies of scale could possibly reduce the premiums to \$0.30/bbl.

Nevertheless, in view of the high premium cost evoked by desulfurization, an extensive research and development effort is under way to develop new processes for oil desulfurizing. Since most refineries have hydrocracking facilities, their adaptation to residual oil treating offers a quickly realizable and perhaps economical desulfurizing technique. Hydrodesulfurization, in which the atmospheric bottoms from a still are fed to a vacuum distillation unit, offers the potential for producing a residual fuel oil product of about 0.5% sulfur. A residual fuel oil product of about 1% sulfur can then be made by blending the high-sulfur vacuum bottoms with low-sulfur gas oil.

A number of other processes involving the combination of various refining processes, such as visbreaking, vacuum distillation, and hydrodesulfurization have also been suggested. Universal Oil Co.'s Isomax Process can be operated as a hydrodesulfurization unit. A large plant is planned for construction in Kuwait. In this plant the sulfur content will be reduced from 5.5 to less than 1%.

Oil-fueled plants will also be faced with some siting problems. Oil storage tanks require considerable space. Oil spills create objectionable conditions. Also, oil-fueled plants often are the source of objectionable soot and other particulate matter.

It might, therefore, be concluded that the pressures that will tend to limit the availability of fuel oil for power generation will be quite significant in the future and the combined problems of availability of a low-sulfur oil, import quotas, and the cost of desulfurizing will tend to limit its use. Oil's share of the electric power generating market will decline quite significantly, but, because of the large increase in power generation, the total demand for oil will increase by approximately one-third.

4.5. Natural Gas

Natural gas has been a source of raw energy for electric power generation in the United States for many years. In the past decade its share of this market has increased approximately 23%, a growth rate that can be attributed to a considerable degree to a disproportionately high degree of industrial development in the areas served by gas. While gas has long possessed the advantages of being a clean and easily handled fuel, it was not until recent years with the advent of increased concern about air pollution that these advantages became a clearly identifiable economic factor. Prior to that, gas competed for the electric power generating load almost entirely on the basis of cost per million Btu. Power plant operators usually took into account the cost penalties incurred by fuel handling, ash disposal, and oil storage and pumping, but these were of limited economic value in selling gas.

To meet market price demands gas is often sold, particularly to power plant and large industrial users, on an interruptible basis. Normally, this means that the purchaser will receive gas at the option of the seller. The purchaser is given adequate warning before his service is interrupted so that he can switch to standby fuels. This affords the electric power plant operator the advantage of using a low-cost, clean-burning fuel at times when it is readily available. This type of service also performs a valuable function for the gas transmission pipeline and distribution companies. Pipeline networks must be sized to meet the demands of the firm gas supply customers — that is, the residential, commercial, and noninterruptible industrial customers. It is essential that pipeline service be available for them whenever they should call for it. However, since the peak demand normally occurs during the period of coldest weather in the wintertime and is somewhat less than that throughout the rest of the year, this could lead to a low load factor and, therefore, higher cost operation. However, through the sale of interruptible gas the pipeline operator is able to improve his load factor, and, by charging the interruptible customer not only the full commodity cost of the gas but also a portion of the fixed charges that accrue to the pipeline, is able to reduce the cost of gas service to the firm customer.

The anticipated rate of growth for a firm gas demand coupled with growth in the pollution-control-stimulated market will place heavy demands upon the indigenous natural gas resources of the United States over the next 3 decades. Alternative sources of supply, such as gas manufactured from coal or imported LNG, have been suggested as possible sources of supply to supplement the dwindling domestic sources.

Overland importation of gas by pipeline from Canada or Mexico offers another supply alternative. Geologically, Mexico appears to have limited possibilities as a source of supply, except for some possible developments in its far southern areas. Canada has tremendous reserves of natural gas, but has a philosophy of rigid export control,lest Canada run short at some future date. With the current oil activity in the Arctic region an apparent indicator of vast oil and gas resources to be discovered, this policy may be eased somewhat. However, the numerous projections made indicate that Canadian sources will not be able to supply much more than 1.5-2 trillion CF of natural gas to the energy economy of the United States.

Research on methods for reducing pollution from central stations using gas has been very limited because natural gas evokes much less of a problem than the other fossil fuels. However, development work on combined cycles using gas turbines and hybrid power plant concepts in which natural gas can be used to complement nuclear power generation offers the possibility of improving gas-fueled power plant cycle efficiencies and thus reducing the quantities of stack gases emitted per unit of power production.

The one advanced power cycle using natural gas that can make a significant contribution to not only reduction of air pollution but also to conservation of natural resources is the fuel cell. This high-efficiency energy conversion unit is the subject of intensive investigation under a program cosponsored by a group of gas pipeline utility companies and a major industrial organization. The fuel cell has the promise of converting natural gas to electricity at efficiencies approaching 50% onsite, thus eliminating electric transmission and distribution losses. Because of the higher efficiency, stack gas production per kWhr of electricity produced is greatly reduced. The value of this technique in conserving the the nation's energy resources has been pointed out in a study[8] conducted by the United States Bureau of Mines. A number of contingency models were developed to show the amount of input energy required to meet the nation's energy demands under various assumptions of an all-gas, an all-electric, and a conventional energy economy. The report concluded with the comment:

> "From the point of view of the efficiency of utilization of natural resources, it might very well be that the development of such a technology [gas-fueled fuel cells] could give us an all-gas economy based upon conversion of other hydrocarbons to gaseous forms to permit their utilization in the fuel cells."[8]

In spite of the possibilities that natural gas offers as a fuel for electric power plant use, it does have several problems. First of all, fuel costs are a problem, although the current cost of pollution control enhances the economic position of gas. Secondly, the current tight domestic supply situation certainly has some impact on the planning of power plant builders. When making the decision as to the plant fuel to be used in a new plant, they must take into account the long-range supply picture. For example, mid-1968 plans for power plants start-ups in 1968-1969-1970 encompassed 60,818 MW of new capacity. Of this only 3886 MW or 6.4% would be solely gas-fueled and another 12.5% would be equipped to use gas as a standby fuel.

Indiscriminate use of pipeline-delivered natural gas to alleviate air pollution problems could also lead to some very severe load-balancing problems for the pipeline operator. For example, the Federal Power Commission, Bureau of Natural Gas, recently conducted a study which estimated that the replacement of all sulfur-bearing fuels consumed in New York City would require about 500 billion CF of natural gas per year.[5] However, the peak demand would be between 2 and 2-1/2 billion CF/day. Therefore, to meet the peaks, additional capacity capable of delivering from 45 to 80% more than the additional annual needs would be required.

4.6. Imported LNG

Since imported liquefied natural gas would affect the national fuel economy, it would have some impact on all of the other fuel forms — coal, oil, domestic natural gas, and nuclear. Therefore, all of the fuel competitors would be concerned. Importation of LNG would increase dependence of our nation's energy economy on foreign resources and would raise Government concerns. At the same time, since it offers the opportunity to advance the industrial development of emerging nations, it merits some attention from international control and financial agencies.

The impact of imported LNG on capital investment trends is fairly clear. Gas-fueled electric generating plants have a clear-cut, first-cost advantage over other fossil-fueled and nuclear plants. This is offset only in part by the incremental investment required for plant site, LNG storage, and vaporization equipment. However, a much larger investment in the fuel supply system is required. By far the major proportion of the investment in an LNG fuel supply system is in the liquefaction plant and tankers. Most of this is invested outside the United States, and the capital would be attracted from those money sources that, in the past, have financed overseas investments of a similar type.

The impact on foreign trade, although not great, would be significant. In relation to the current levels of imports of approximately $25 billion/year, the importation of approximately $300 million worth of LNG would be but a small factor. However, taken in relation to net energy imports, this amount of LNG could have some significance. For example, in 1965, total net imports of energy were 4.074 X 10^{15} Btu. Crude oil and petroleum products totaled 4.97 X 10^{15} Btu, and natural gas 0.437 x 10^{15} Btu. This was counterbalanced, in part, by the export of 1.31 X 10^{15} Btu as bituminous coal and lignite. The proposed LNG imports for power generation would add about 18% to the total imported energy.

Negotiations for the necessary large quantities of gas would probably affect trade relations between the source country and the United States. Since the gas is largely controlled by the government of the country in which it is located, negotiations must be carried out through a state agency. In the past, these negotiations have largely been conducted by representatives of potential importers and the ministries of industrial development or petroleum.

The Federal Power Commission now regulates the import or export of natural gas. Section 3 of the Natural Gas Act states: "After 6 months from the date on which this Act takes effect, no person shall export any natural gas from the United States to a foreign country or import any natural gas from a foreign country, without first having secured an order of the Commission authorizing it to do so. The Commission shall issue such order upon application, unless, after opportunity for hearing, it finds that the proposed exportation or importation will not be consistent with the public interest." Whether or not this wording would apply to the importation of liquefied natural gas remains to be settled. However, in 1963 the Federal Power Commission announced that it was of the general opinion that it "may have jurisdiction over the import and export of liquefied methane under Section 3 of the Natural Gas Act."[6] The Federal Power Commission has not as yet clarified its position as to the general importation of liquefied natural gas, although discussions with staff personnel indicate there could be conditions under which importation of liquefied natural gas solely for power generation might not be jurisdictional, except for authorization and granting of a Presidential Permit.

The Federal Power Commission is charged with the responsibility of carrying out Executive Order No. 10485, issued in 1953. This order provides that a Presidential Permit "be obtained for the construction and maintenance at the borders of the United States, of facilities for the exportation or importation of natural gas." The Federal Power Commission has the responsibility of receiving applications and issuing appropriate permits. These permits are issued when it is found that issuance of the permit is consistent with the public interest and upon favorable recommendations from the Secretary of State and the Secretary of Defense. The purpose of this permit is to protect the integrity of the United States border rather than to control the importation of gas. Since there are no border facilities interconnected between the countries involved, there is a question as to whether or not a Presidential Permit may really be required for LNG importation.

In addition to the already recognized problems of control of natural gas import and the issuance of Presidential Permits, the possible importation of large quantities of liquefied natural gas for power generation would evoke many as yet unrecognized, or at least unstated, problems. For example, the impact of imported LNG on existing gas rate structures certainly will evoke some regulatory response from not only the Federal Power Commission but perhaps, for wholly intrastate usage, state public utility commissions.

Segregation of imported natural gas, if indeed it is natural gas, may be difficult to show; thus, the commingling theory may raise many questions as to the jurisdictional nature of imported LNG. The fundamental question of whether or not liquefied natural gas is really natural gas, within the meaning of the Natural Gas Act, still remains to be resolved.

Importation of crude oil, finished products, and liquefied gases is presently controlled by the Oil Import Administration. Import quotas, currently at the level of 12.2% of domestic production, have been set to limit the dependence of the U. S. oil economy on foreign sources. Since crude oil and finished products comprise by far the preponderance of energy imports into the United States, the Oil Import Administration regulations do have a significant effect on the energy trade balance of the United States. Under the present definition it appears that liquefied natural gas would not come under control of the Oil Import Administration. Liquefied gases are defined as "ethane, propane, butanes, ethylene, propylene, and butylene (but not methane), which are recovered from natural gas or produced in the refining of petroleum and which, to be maintained in a liquid state at ambient temperatures, must be kept under greater than atmospheric pressures."

Although the exemption is specifically limited to methane, it was the opinion of Department of the Interior personnel that liquefied natural gas, so long as it was constituted principally of methane, would be exempt. The exemption actually appears to be based largely on the use to which the product would be put. If the imported LNG were vaporized and used directly as a fuel, it would not be controlled. However, if it were fractionated after receipt and some of the controlled liquefied gases, such as ethane, propane, and butane, were recovered for other than fuel use, these would immediately become controlled items and import tickets would be required.

Whether this exemption from control can be maintained in the future is problematical. Certainly there will be pressure from domestic producers to exert some sort of control over the importation of LNG. It is likely that this pressure will be felt before the present level of oil imports is reached. If the import allocation of 12.2 % of domestic production were the criterion, import of well over 2 trillion CF of natural a year would be tolerable. It is doubtful, however, that this level would be reached without some outcry from domestic producers.

The impact of imported LNG on the tariff laws is somewhat more difficult to assess. At the present time, there is no tariff on natural gas. However, there is a continuing effort on the part of the Congressional delegations from some of the gas-producing states to impose a tariff. The latest effort, embodied in House Bill H. R. 10233, would impose a tariff of 3¢/1000 CF on all imported natural gas. Since all of the imported natural gas now comes from either Canada or Mexico, what little protectionist sentiment that does prevail is more than counterbalanced by the desire to maintain good trade relations with our continental neighbors. When dealing with South American or African countries, this effect is diminished and the protectionists may have a somewhat greater degree of success in imposing some form of tariff. The duty on crude oil is 10.5¢/bbl for oil over 25 gravity and 5.25¢/bbl for oil under 25 gravity. Duties on gasoline are 1.25¢/gal. On all other products, they are 0.25¢/gal. On an energy content basis, a duty comparable with the 10.5¢/bbl for high-gravity oil would be about 1.75¢/1000 CF.

Since imported LNG will be a commodity in marine traffic, it will be of concern to several other Federal agencies. The Federal Maritime Commission will be concerned as to the application of the Merchant Marine Act 46 USC 883, the so-called Jones Act. The Act says in part, "no merchandise shall be transported by water... between points in the United States, including districts, territories, and possessions thereof... in any other vessel than a vessel built in and documented under the laws of the United States, and owned by persons who are citizens of the United States, or vessels to which the privilege of engaging in coast-wide trade is extended by Section 13 or 808 of this title." Because of the higher cost of construction of American-built ships and the higher operating costs, imposition of the Jones Act restrictions would significantly affect the cost of importing liquefied natural gas from Alaska to West Coast states. Several exemptions have been granted to the Act in the past, notably for the transportation of lumber from the Pacific Northwest to Puerto Rico and various types of commerce between Alaska and the Northwest United States, and between ports along the Great Lakes. However, these were time-limited exemptions granted for specific cases and it appears that a blanket exemption for the interstate transportation of Alaskan LNG may be very difficult to obtain. A specific exemption for the Alaska-Hawaii shipment might be somewhat easier to get because of the limited opposition offered by the labor unions and subsidized shipline on the Alaska-Hawaii run.

Safety in handling and unloading LNG in U. S. ports is, and will continue to be, the concern of the United States Coast Guard. Because of its concern the Coast Guard is considering the funding of investigations of the effects of dumping large quantities of LNG into salt water, as might occur in a collision. The Coast Guard has had experience with at least one importation program, that of the LNG brought into Boston Harbor in the winter of 1968-69. While it was necessary to hold the LNG tanker well out from the dock and transfer the LNG into cryogenic trucks mounted on a barge, the Coast Guard had an opportunity to observe that LNG can be handled safely and with a minimum of danger to surrounding structures. Hopefully, the current work under way on developing and demonstrating safe handling methods of large volumes of LNG will minimize or eliminate the concern of safety officials.

The impact of imported LNG on the industrial development of the importing region is likely to be significantly different on the East and West Coasts. On the East Coast where the pattern of industrial growth is fairly well developed, imported liquefied natural gas would not affect the economics of electric supply sufficiently to cause any great influx of electric-power-consuming industries. Cold utilization processes could spawn new industries, and perhaps improve the economics of industries such as refineries and steelmaking that are already in existence. If importation of LNG for power generation could pave the way for other importation programs that would utilize LNG as a chemical feedstock, there may be some increase in petrochemical production. The most important effect would be the possibilities offered for integration of the power plant fuel import program with base-load import for a gas utility.

However, on the West Coast where the industrial growth pattern is still in the state of flux, imported LNG could have a major impact. For example, in southern California where there is a great demand for both energy and water, the imported LNG could become the part of a large energy complex in which the recovered refrigeration could be used for water desalting or industrial gas production, (LOX and LN_2) while the exhaust heat from the electric power generating plant steam turbines could be used for water desalting. In concept this would be quite similar to the nuclear-energy-based plant proposed for the Bolsa Island project.

LNG importation programs, even of the size visualized in this study, will have little impact on changing employment patterns. Most of the labor requirements will be met by foreign employees, either in the foreign liquefaction plant site or as foreign crews on the tankers. Little labor is involved at the receiving terminal. The labor skills that will be required are quite similar to those now necessary for electric power plant operation. However, in the combination facility where a number of products would be produced, a broader employment base might be developed. An analysis of the labor requirements for such a facility is beyond the scope of this present study.

Imported LNG could have a significant impact on changing public health and safety regulations. Its availability as a low-cost alternative to high-sulfur fuels could accelerate the application of stringent sulfur control regulations. The Air Quality Act of 1967 recognizes that time is needed to develop adequate pollution control technology. If importation of LNG is a technically and economically feasible alternative, it may be worthwhile to devote some portion of the air pollution control research funds to improving the methods of liquefying, transporting, and utilizing LNG. At the same time, the availability of this alternative may stimulate the domestic suppliers of high-sulfur fuels, in particular the coal industry, to increase their research efforts in developing sulfur-control methods.

Safety in the handling and storage of LNG will undoubtedly be the subject of some concern to the potential user. Effective codes now exist for the construction of LNG plants on utility plant sites. The National Fire Protection Association Code 59A, which sets clearance limits, outlines methods of construction, etc., is constantly being reviewed and revised by a group of competent industry leaders. Numerous programs on safety research are being conducted by the LNG industry. A program currently under way at the Institute of Gas Technology will document and demonstrate safe handling and storage techniques for LNG. The American Gas Association has, for the past several years, been sponsoring research on LNG safety, and numerous industrial organizations have conducted their own programs to improve the safety of LNG.

4.7. Impact of Imported LNG on Nuclear Power Industry

What effect could imported LNG have on the nuclear power industry? Prior to cost escalations of the past two years, nuclear power costs for investor-owned utilities were most commonly estimated to be in a general range of 4-5 mills/kWhr. (This range applicable to plants on order, extended upward to 6 or 7 mills/kWhr for some domestic locations and downward to 3.3-3.5 mills/kWhr for publicly-owned facilities such as TVA plants.) Actually, no nuclear plants have yet demonstrated operation at the 5-mill cost level. If they had, little competition could be offered, on a straight cost basis, by LNG-based plants having fuel costs alone of 4 mills/kWhr at best.

Even assuming major upward changes in future nuclear power costs, will LNG ever generate lower cost electricity? Probably not, but the premium one would have to pay for a clean-fueled plant close to metropolitan centers might, in the future, be much less than now appears to be the case.

The two major cost elements, fuel costs and operating costs, are entirely different for a nuclear or gas-fueled plant. There is little doubt that nuclear fuel costs will be less than imported LNG. In 1967 an EEI study group prepared a report that indicated that nuclear fuel cycle costs would be approximately 1.8 mills/kWhr in 1970 and even lower in later years. On a strict Btu comparison basis of alternative fuels, imported LNG would have to be available at 19¢/million Btu, an impossible target. Even if nuclear raw materials' costs increase drastically to the $15/lb level, fuel cycle costs would increase to only about 2.2 mills/kWhr, and the LNG break-even cost to about 23¢/million Btu.

But the second major cost element greatly favors gas-fueld plants over nuclear plants. As interest rates rise and plant load factors decline, the fixed charges will increase. Early nuclear plants will be base-loaded, just as it was assumed the LNG-fueled plants will be. But as these plants grow in number, the load factor must decline toward the average utility generation load factor of something less than 60%. The effects of higher money costs, lower load factors, and higher investment costs are clearly shown in Figure 4.7.-1.

As can be seen, in all cases, the LNG-fueled plant has lower fixed charges than the nuclear plant. The question is, are the fixed cost savings sufficient to offset the higher fuel cycle charges. As a close approximation, it can be estimated that 1 mill/kWhr in fixed charges is equivalent to 10.4¢/million Btu in the cost of fuel. Therefore, by equating the differentials in fixed charges shown in Figure 4.7.-1 to fuel costs, the break-even cost of LNG fuel can be estimated.

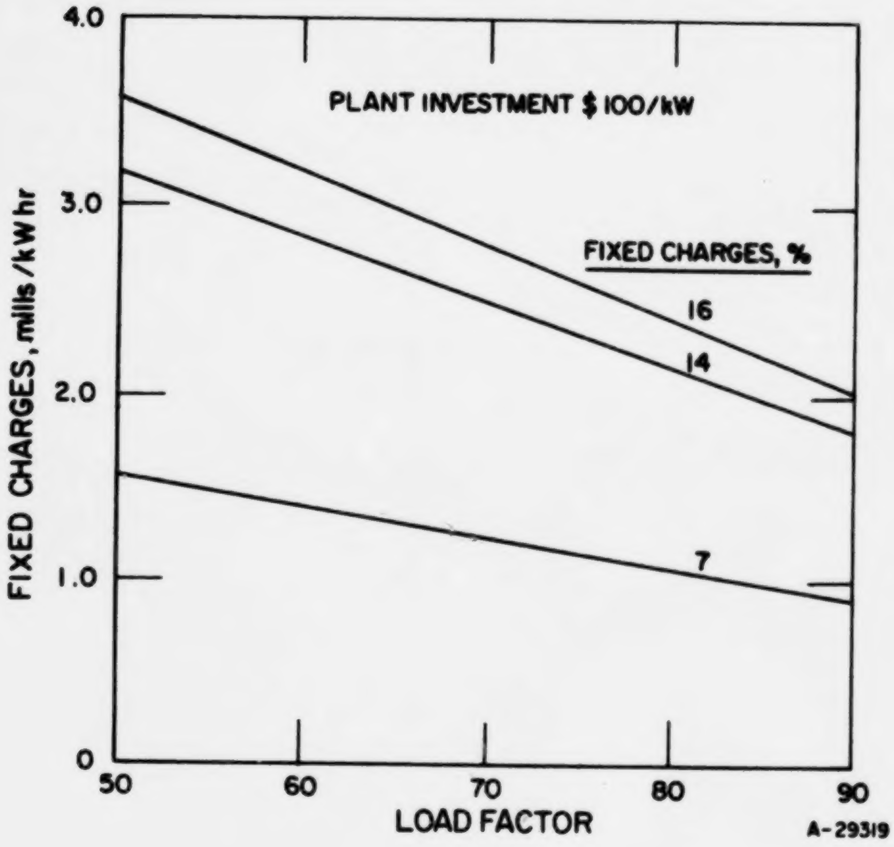

Figure 4. 7. -1. EFFECT OF LOAD FACTOR AND FIXED RATE AT FIXED CHARGES

The break-even cost for LNG can be computed by the formula:

$$\text{Break-Even Cost} = \frac{10^6}{H}\left[\frac{(I_n - I_g)(r)}{8.76\ \text{L. F.}} + F_n\right]$$

where

F_n = nuclear plant fuel cost, mills/kWhr

H = LNG-fueled plant heat rate, Btu/kWhr

I_n = nuclear plant investment cost, \$/kW

I_g = LNG-fueled plant investment, \$/kW

L. F. = generation plant load factor, decimal

r = fixed charge rate, decimal

For example: assuming a fixed charge rate of 14%, a 0.60 load factor, nuclear plant investment of \$175/kW, an LNG-fueled plant investment of \$100/kW, and nuclear plant fuel costs of 2 mills/kWhr, the break-even cost is —

$$C = \frac{10^6}{9250}\left[\left(\frac{14}{876 \times 0.6}\right)(175 - 100) + 2.0\right] = 43.2¢/\text{million Btu.}$$

As can be seen, the factors that affect the break-even cost are fixed charge rates, load factors, investment for both nuclear and gas-fueled plants, and the cost of nuclear fuel. While the present 14% rate is adequate for making comparative evaluations, the strong upward pressures on the cost of capital make the possibility of significantly higher fixed charge rates much more likely. A rate of 16% might not be at all unrealistic in the next decade; this factor alone would increase the break-even cost of LNG by approximately 2.5¢/million Btu. It is likely that the load factor will ultimately settle in the range of 0.6, although, if it were 0.54, the all-industry average in 1966, the break-even cost would be about 2.1¢/million Btu higher.

Nuclear fuel costs, may also vary from present projections, forces are at work that might either increase or decrease these costs. As competition tightens, uranium oxide prices will increase. Since they contribute approximately 25% of the final fuel cost, an increase to $15/lb of U_3O_8 could create a significant cost difference. However, at the same time, performance improvements, and lower cost fabrication and recovery techniques may hold fuel costs reasonably well in line.

One item that has not been taken into account in this computation of break-even costs is the differential in operating costs. While this is small for both gas-fueled and nuclear plants, it will always be somewhat higher for nuclear plants because of their need for more and better trained personnel and special safety requirements. Although not used in computing break-even costs, it might be anticipated that operating costs would be about 0.2 mill/kWhr higher for nuclear plants, thus increasing the break-even cost by about 2.2¢/million Btu.

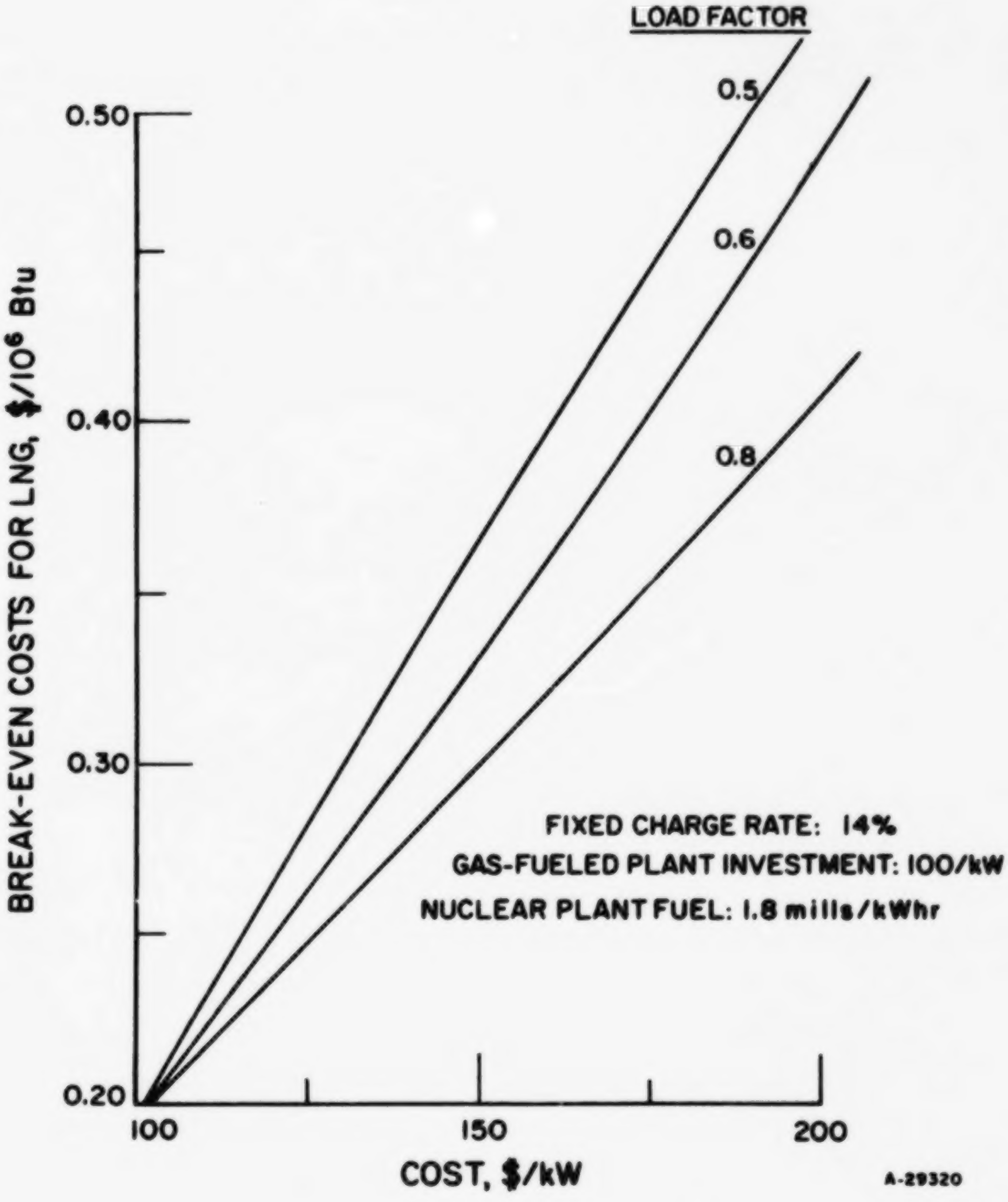

Figure 4. 7. -2. NUCLEAR POWER PLANT INVESTMENT

With all of these variable costs affecting the break-even price of LNG, it would be easy to select a set of credible circumstances that would indicate that imported LNG could produce electricity at lower costs than nuclear energy.

However, this will probably not be the case. The tremendous research effort being expended on nuclear power research will slow the rate of increase of nuclear plant costs and very probably also will reduce fuel cycle costs. The inherent economic burden that imported LNG must carry can be only partially offset by continued research to reduce costs through equipment improvements and to improve recovery of the refrigeration energy. Nevertheless, it is abundantly clear that imported LNG can achieve a position much more nearly competitive than that indicated by a cursory examination of the published costs of bus bar energy from nuclear power plants and imported LNG

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Section 5.0. Conclusions and Recommendations

On the basis of our analysis we have come to the following generalized conclusions:

1. There is adequate natural gas available at several locations in the world from which it might be economically supplied to the coastal regions of the United States.
2. As is evidenced by the imminent use of ocean transported LNG for power generation in Japan before the end of 1969, the concept of liquefying natural gas in an area of energy surplus, transporting it by marine tanker to an area of energy deficiency, and burning it in an electric power plant is technically feasible.
3. If natural gas can be purchased in the source country at "reasonable" prices, it can be liquefied and transported from Venezuela and Libya to East Coast centers and from Western Canada to West Coast centers at a cost which would be competitive with high-sulfur fuels when the cost of desulfurizing is considered.
4. Nuclear fuel electric power plants, incurring no sulfur control penalties, would appear to be more economical in both coastal regions, provided safety problems do not impose an undue cost penalty and the current upward trend in nuclear power plant costs can be stemmed.
5. LNG-based power generating plants would have no greater safety, aesthetic, thermal pollution, or other environmental problems than similarly sized fossil-fueled plants. They would offer a significant improvement in aesthetic benefits through the elimination of coal storage areas and fuel oil tank farms, and possibly through improved safety as compared to nuclear plants.
6. LNG technology is quite new and still evolving rapidly. Technological trends to larger, simpler, and lower cost equipment indicate a further improvement in LNG's competitive position. In particular, significant improvements could be made through the development of better cold utilization technology.
7. There may be several noneconomic factors that would slow the acceptance of LNG as a fuel for power generation. Among these are the uncertainties of regulatory status, the possible impact on the domestic natural gas and pipelining industry, political problems concerned with developing trade with foreign countries, and a reluctance on the part of the electric industry to rely on a foreign energy source.
8. Industrial gas users also can use imported LNG, thus further reducing air pollution.

We recommend that the NAPCA develop a program that will stimulate the acceptance of imported LNG as a low-sulfur fuel for power generation.

We recommend that a broad-based investigation of the technical, economic, regulatory, and political factors that might slow its acceptance be undertaken. This program would seek to instruct control agencies, regulators, suppliers, users, investors, and the general public of the merits of this fuel as a means of reducing air pollution.

We recommend a program that would contain, but not be limited to, the following elements:

1. Definition of both nuclear and low-sulfur fuel competition to imported LNG.
2. Optimization of LNG-fueled power plant cycles to utilize available refrigeration to reduce costs, increase capacity, etc.
3. Analysis of use of LNG in nonconventional power cycles such as MHD, etc.
4. Determination of feasibility of imported LNG for noncoastal locations.
5. A detailed assessment of the effect of LNG importation on the domestic natural gas industry.
6. Evaluation of the potential use of imported LNG as an industrial boiler fuel.

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APPENDIX B. Data

Conversion Tables for Liquefied Natural Gas –

The following tables[1] present various conversion factors for natural gas of selected molecular weights.

CONVERSION TABLE FOR NATURAL GAS - 16.0 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	23.748	23.673	23.619	23.155
1 TON (2000.0LB)	47495.6	47346.9	47237.6	46310.1
1 TON (2204.6LB)	52354.3	52190.5	52070.0	51047.7
1 TON (2240.0LB)	53195.0	53028.5	52906.1	51867.4
1 GALLON	84.146	83.883	83.689	82.046
1 BARREL	3534.1	3523.1	3514.9	3445.9

CONVERSION CONSTANTS

DENSITY = 26.506 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
629.454	627.483	626.035	613.743

CONVERSION TABLE FOR NATURAL GAS - 16.5 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	23.024	22.951	22.898	22.449
1 TON (2000.0LB)	46047.0	45902.9	45797.0	44897.8
1 TON (2204.6LB)	50757.7	50598.8	50482.0	49490.8
1 TON (2240.0LB)	51572.7	51411.3	51292.6	50285.5
1 GALLON	83.499	83.238	83.046	81.415
1 BARREL	3507.0	3496.0	3487.9	3419.4

CONVERSION CONSTANTS

DENSITY = 27.129 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
624.616	622.661	621.223	609.026

CONVERSION TABLE FOR NATURAL GAS - 17.0 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	22.342	22.272	22.221	21.784
1 TON (2000.0LB)	44683.8	44543.9	44441.1	43568.5
1 TON (2204.6LB)	49254.9	49100.8	48987.4	48025.6
1 TON (2240.0LB)	50045.8	49889.2	49774.0	48796.8
1 GALLON	83.076	82.816	82.625	81.003
1 BARREL	3489.2	3478.3	3470.2	3402.1

CONVERSION CONSTANTS

DENSITY = 27.815 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
621.451	619.505	618.075	605.940

CONVERSION TABLE FOR NATURAL GAS - 17.5 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	21.697	21.629	21.579	21.155
1 TON (2000.0LB)	43394.0	43258.2	43158.4	42311.0
1 TON (2204.6LB)	47833.2	47683.5	47573.5	46639.4
1 TON (2240.0LB)	48601.3	48449.2	48337.4	47388.3
1 GALLON	82.487	82.229	82.039	80.428
1 BARREL	3464.5	3453.6	3445.6	3378.0

CONVERSION CONSTANTS

DENSITY = 28.439 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
617.045	615.113	613.694	601.644

CONVERSION TABLE FOR NATURAL GAS - 18.0 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	21.090	21.024	20.976	20.564
1 TON (2000.0LB)	42180.2	42048.2	41951.1	41127.5
1 TON (2204.6LB)	46495.2	46349.7	46242.7	45334.8
1 TON (2240.0LB)	47241.8	47094.0	46985.3	46062.7
1 GALLON	81.938	81.682	81.493	79.893
1 BARREL	3441.4	3430.6	3422.7	3355.5

CONVERSION CONSTANTS

DENSITY = 29.063 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
612.938	611.019	609.609	597.640

CONVERSION TABLE FOR NATURAL GAS - 19.0 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	19.972	19.910	19.864	19.474
1 TON (2000.0LB)	39944.2	39819.1	39727.2	38947.2
1 TON (2204.6LB)	44030.4	43892.6	43791.3	42931.5
1 TON (2240.0LB)	44737.4	44597.4	44494.5	43620.9
1 GALLON	80.925	80.671	80.485	78.905
1 BARREL	3398.8	3388.2	3380.4	3314.0

CONVERSION CONSTANTS

DENSITY = 30.310 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
605.357	603.462	602.069	590.248

CONVERSION TABLE FOR NATURAL GAS - 20.0 M.W.

	CUBIC FEET OF NATURAL GAS AT 60 F.			
LIQUID AT -260 F.	14.650PSI	14.696PSI	14.730PSI	15.025PSI
1 POUND	18.966	18.906	18.863	18.492
1 TON (2000.0LB)	37931.7	37813.0	37725.7	36985.0
1 TON (2204.6LB)	41812.1	41681.2	41585.0	40768.6
1 TON (2240.0LB)	42483.5	42350.5	42252.8	41423.2
1 GALLON	79.852	79.602	79.418	77.859
1 BARREL	3353.8	3343.3	3335.6	3270.1

CONVERSION CONSTANTS

DENSITY = 31.495 LB/CU.FT.

1 GALLON = 0.133681 CU.FT.

1 BARREL = 42 GALLONS

RATIOS OF CU.FT. OF GAS TO CU.FT. OF LIQUID

14.650PSI	14.696PSI	14.730PSI	15.025PSI
597.332	595.462	594.088	582.423

Imported LNG Cost Computations

A cost summary for imported LNG is provided for each alternative overseas gas source evaluated and at each assumed foreign gas purchase price. The various items on the printout sheet are:

1. Location and price of gas at liquefaction plants.
2. Escalation factors for plant and transportation operating cost, percent per year.
3. Investment for each part of the system.
4. Total cost of all gas purchased, including that required for liquefaction plant fuel, tanker boil-off, and losses.
5. Fixed and operating costs of natural gas liquefaction plant. Unit cost includes cost of purchased gas plus liquefaction costs.
6. Fixed and operating costs of LNG tankers. Unit cost includes f.o.b. cost of LNG delivered to tanker plus transportation cost.
7. Fixed and operating costs of receiving terminal. Unit cost includes cost of LNG delivered over ship's rail.
8. Average cost of gas over the life of the project.

VENEZUELAN GAS AT 0.10/MMBTU[1] 1-10-69 PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR 3.000 PERCENT } [2]
TANKER ESCALATION FACTOR 2.000 PERCENT }
RECEIVING TERMINAL ESCALATION FACTOR 3.000 PERCENT }

TOTAL PROJECT INVESTMENT[3]	
LIQUEFACTION PLANT	545.000
OCEAN TRANSPORTATION	330.000
RECEIVING TERMINAL	88.000
TOTAL INVESTMENT	963.000

LIQUEFACTION PLANT CAPACITY 2500.000 MILLION CUBIC FEET PER DAY
8 648.200. THOUSAND BARREL TANKERS

GAS INPUT/OUTPUT RATIO 1.235

PROJECT LIFE 20 YEARS GAS DELIVERED 2030.400 BILLION BTU/DAY 741.096 MILLION MMBTU/YEAR

TOTAL PURCHASED GAS[4]		LIQUEFACTION PLANT[5]			OCEAN TRANSPORTATION[6]			RECEIVING TERMINAL		TOTAL COST	UNIT COST[7]
UNIT COST CENTS PER MMBTU	TOTAL COST MILLIONS OF DOLLARS	FIXED COST MILLIONS	VARIABLE COSTS OF DOLLARS	F.O.B. COST CENTS PER MMBTU	FIXED COST MILLIONS	VARIABLE COST OF DOLLARS	C.I.F. COST CENTS PER MMBTU	FIXED COST MILLIONS	VARIABLE COST OF DOLLARS	MILLIONS OF DOLLARS	OF GAS CENTS PER MMBTU
10.000	91.530	109.000	54.500	34.412	33.000	27.628	42.593	12.869	8.780	337.308	45.514
10.000	91.530	109.000	55.317	34.522	33.000	27.850	42.733	12.209	8.911	337.819	45.583
10.000	91.530	109.000	56.159	34.636	33.000	28.077	42.877	11.549	9.046	338.363	45.657
10.000	91.530	109.000	57.026	34.753	33.000	28.309	43.026	10.889	9.186	338.942	45.735
10.000	91.530	109.000	57.920	34.874	33.000	28.545	43.178	10.229	9.329	339.555	45.817
10.000	91.530	109.000	58.840	34.998	33.000	28.786	43.335	9.569	9.477	340.203	45.905
10.000	91.530	109.000	59.787	35.126	33.000	29.031	43.496	8.909	9.629	340.889	45.998
10.000	91.530	109.000	60.764	35.257	33.000	29.282	43.661	8.249	9.786	341.613	46.095
10.000	91.530	109.000	61.769	35.393	33.000	29.538	43.832	7.589	9.948	342.376	46.198
10.000	91.530	109.000	62.805	35.533	33.000	29.798	44.006	6.929	10.114	343.178	46.306
10.000	91.530	109.000	63.871	35.677	33.000	30.064	44.186	6.269	10.286	344.023	46.420
10.000	91.530	109.000	64.970	35.825	33.000	30.336	44.371	5.609	10.462	344.909	46.540
10.000	91.530	109.000	66.101	35.978	33.000	30.612	44.561	4.949	10.644	345.839	46.665
10.000	91.530	109.000	67.267	36.135	33.000	30.895	44.757	4.289	10.832	346.814	46.797
10.000	91.530	109.000	68.467	36.297	33.000	31.183	44.957	3.629	11.025	347.836	46.935
10.000	91.530	109.000	69.704	36.464	33.000	31.476	45.164	2.969	11.223	348.905	47.079
10.000	91.530	109.000	70.978	36.636	33.000	31.776	45.376	2.309	11.428	350.023	47.230
10.000	91.530	109.000	72.289	36.813	33.000	32.081	45.594	1.649	11.639	351.191	47.388
10.000	91.530	109.000	73.641	36.995	33.000	32.393	45.819	0.989	11.856	352.411	47.552
10.000	91.530	109.000	75.032	37.183	33.000	32.711	46.049	0.329	12.080	353.684	47.724

AVERAGE COST OF GAS[8] 46.457 CENTS PER MMBTU LEVELIZED COST OF GAS 46.136 CENTS PER MMBTU

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    VENEZUELAN GAS  AT  0.05/MMBTU      1-10-69   PROJ. 8926

 LIQUEFACTION COST ESCALATION FACTOR       3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

  TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      545.000         LIQUEFACTION PLANT CAPACITY 2500.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    330.000            8     648.200  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       88.000
TOTAL INVESTMENT        963.000         GAS INPUT/OUTPUT RATIO        1.235

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY           741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT              OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED     VARIABLE  MILLIONS    OF GAS
CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST      COST    CENTS PER    COST      COST       OF      CENTS PER
 MMBTU     OF DOLLARS MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS     MMBTU

  5.000      45.765   109.000   54.500     28.237   33.000     27.628     36.418   12.869     8.780   291.542     39.339
  5.000      45.765   109.000   55.317     28.347   33.000     27.850     36.558   12.209     8.911   292.054     39.408
  5.000      45.765   109.000   56.159     28.461   33.000     28.077     36.702   11.549     9.046   292.598     39.481
  5.000      45.765   109.000   57.026     28.578   33.000     28.309     36.850   10.889     9.186   293.177     39.559
  5.000      45.765   109.000   57.920     28.698   33.000     28.545     37.003   10.229     9.329   293.790     39.642
  5.000      45.765   109.000   58.840     28.822   33.000     28.786     37.160    9.569     9.477   294.439     39.730
  5.000      45.765   109.000   59.787     28.950   33.000     29.031     37.321    8.909     9.629   295.124     39.822
  5.000      45.765   109.000   60.764     29.082   33.000     29.282     37.486    8.249     9.786   295.848     39.920
  5.000      45.765   109.000   61.769     29.218   33.000     29.538     37.656    7.589     9.948   296.611     40.023
  5.000      45.765   109.000   62.805     29.357   33.000     29.798     37.831    6.929    10.114   297.413     40.131
  5.000      45.765   109.000   63.871     29.501   33.000     30.064     38.011    6.269    10.286   298.257     40.245
  5.000      45.765   109.000   64.970     29.650   33.000     30.336     38.196    5.609    10.462   299.144     40.365
  5.000      45.765   109.000   66.101     29.802   33.000     30.612     38.386    4.949    10.644   300.074     40.490
  5.000      45.765   109.000   67.267     29.960   33.000     30.895     38.581    4.289    10.832   301.049     40.622
  5.000      45.765   109.000   68.467     30.122   33.000     31.183     38.782    3.629    11.025   302.071     40.760
  5.000      45.765   109.000   69.704     30.288   33.000     31.476     38.989    2.969    11.223   303.140     40.904
  5.000      45.765   109.000   70.978     30.460   33.000     31.776     39.201    2.309    11.428   304.257     41.055
  5.000      45.765   109.000   72.289     30.637   33.000     32.081     39.419    1.649    11.639   305.426     41.212
  5.000      45.765   109.000   73.641     30.820   33.000     32.393     39.643    0.989    11.856   306.646     41.377
  5.000      45.765   109.000   75.032     31.007   33.000     32.711     39.874    0.329    12.080   307.919     41.549

AVERAGE COST OF GAS    40.282  CENTS PER MMBTU               LEVELIZED COST OF GAS    39.960  CENTS PER MMBTU
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    ALGERIA GAS  AT  0.08/MMBTU   1-10-69    PROJ. 8926

  LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

  TANKER ESCALATION FACTOR                   2.000  PERCENT

  RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

   TOTAL PROJECT INVESTMENT
  LIQUEFACTION PLANT       510.000        LIQUEFACTION PLANT CAPACITY 2320.000 MILLION CUBIC FEET PER DAY
  OCEAN TRANSPORTATION     650.000          16    639.750  THOUSAND BARREL TANKERS
  RECEIVING TERMINAL        88.800
  TOTAL INVESTMENT        1248.800        GAS INPUT/OUTPUT RATIO       1.296

  PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

   TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
  UNIT COST TOTAL COST  FIXED      VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED     VARIABLE  MILLIONS   OF GAS
  CENTS PER   MILLIONS  COST       COSTS     CENTS PER   COST       COST    CENTS PER    COST       COST        OF     CENTS PER
   MMBTU    OF DOLLARS MILLIONS OF DOLLARS  MMBTU      MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS  DOLLARS    MMBTU

    8.000     76.856   102.000    51.000     31.015   65.000    53.316    46.980   12.986     8.845    370.004    49.926
    8.000     76.856   102.000    51.765     31.118   65.000    53.732    47.140   12.320     8.978    370.652    50.014
    8.000     76.856   102.000    52.552     31.225   65.000    54.156    47.303   11.654     9.114    371.335    50.106
    8.000     76.856   102.000    53.364     31.334   65.000    54.590    47.471   10.988     9.254    372.054    50.203
    8.000     76.856   102.000    54.200     31.447   65.000    55.031    47.644   10.322     9.398    372.810    50.305
    8.000     76.856   102.000    55.061     31.563   65.000    55.482    47.821    9.656     9.547    373.604    50.412
    8.000     76.856   102.000    55.948     31.683   65.000    55.942    48.002    8.990 -   9.700    374.438    50.524
    8.000     76.856   102.000    56.861     31.806   65.000    56.411    48.189    8.324     9.858    375.312    50.642
    8.000     76.856   102.000    57.802     31.935   65.000    56.889    48.380    7.658    10.021    376.228    50.766
    8.000     76.856   102.000    58.771     32.064   65.000    57.377    48.577    6.992    10.188    377.186    50.895
    8.000     76.856   102.000    59.769     32.199 · 65.000    57.874    48.779    6.326    10.361    378.188    51.030
    8.000     76.856   102.000    60.797     32.337   65.000    58.382    48.986    5.660    10.538    379.235    51.172
    8.000     76.856   102.000    61.856     32.480   65.000    58.899    49.199    4.994    10.721    380.329    51.319
    8.000     76.856   102.000    62.947     32.627   65.000    59.427    49.417    4.328    10.910    381.470    51.473
    8.000     76.856   102.000    64.070     32.779   65.000    59.966    49.641    3.662    11.104    382.660    51.634
    8.000     76.856   102.000    65.228     32.935   65.000    60.515    49.872    2.996    11.304    383.901    51.801
    8.000     76.856   102.000    66.419     33.096   65.000    61.075    50.108    2.330    11.510    385.143    51.976
    8.000     76.856   102.000    67.647     33.262   65.000    61.647    50.351    1.664    11.722    386.538    52.157
    8.000     76.856   102.000    68.911     33.432   65.000    62.230    50.600    0.998    11.940    387.938    52.346
    8.000     76.856   102.000    70.214     33.608   65.000    62.824    50.856    0.332    12.165    389.394    52.542

   AVERAGE COST OF GAS    51.062  CENTS PER MMBTU                 LEVELIZED COST OF GAS    50.680  CENTS PER MMBTU
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B-8

8926

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   ALGERIA GAS  AT  0.06/MMBTU   1-10-69    PROJ. 8926
LIQUEFACTION COST ESCALATION FACTOR       3.000  PERCENT

TANKER ESCALATION FACTOR                  2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR      3.000  PERCENT
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 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      510.000          LIQUEFACTION PLANT CAPACITY 2320.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    650.000            16    639.750  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       88.800
TOTAL INVESTMENT       1248.800          GAS INPUT/OUTPUT RATIO       1.296
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PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS   OF GAS
CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST      COST     CENTS PER    COST      COST       OF     CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU     MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS     MMBTU

  6.000     57.641   102.000   51.000      28.423   65.000     53.316      44.388   12.986      8.845   350.790     47.334
  6.000     57.641   102.000   51.765      28.526   65.000     53.732      44.547   12.320      8.978   351.438     47.421
  6.000     57.641   102.000   52.552      28.632   65.000     54.156      44.711   11.654      9.114   352.121     47.513
  6.000     57.641   102.000   53.364      28.742   65.000     54.590      44.879   10.988      9.254   352.840     47.610
  6.000     57.641   102.000   54.200      28.854   65.000     55.031      45.051   10.322      9.398   353.596     47.712
  6.000     57.641   102.000   55.061      28.971   65.000     55.482      45.228    9.656      9.547   354.390     47.819
  6.000     57.641   102.000   55.948      29.090   65.000     55.942      45.410    8.990      9.700   355.224     47.932
  6.000     57.641   102.000   56.861      29.213   65.000     56.411      45.596    8.324      9.858   356.098     48.050
  6.000     57.641   102.000   57.802      29.340   65.000     56.889      45.788    7.658     10.021   357.014     48.173
  6.000     57.641   102.000   58.771      29.471   65.000     57.377      45.984    6.992     10.188   357.972     48.303
  6.000     57.641   102.000   59.769      29.606   65.000     57.874      46.186    6.326     10.361   358.974     48.438
  6.000     57.641   102.000   60.797      29.745   65.000     58.382      46.393    5.660     10.538   360.021     48.579
  6.000     57.641   102.000   61.856      29.888   65.000     58.899      46.606    4.994     10.721   361.115     48.727
  6.000     57.641   102.000   62.947      30.035   65.000     59.427      46.824    4.328     10.910   362.256     48.881
  6.000     57.641   102.000   64.070      30.186   65.000     59.966      47.049    3.662     11.104   363.446     49.041
  6.000     57.641   102.000   65.228      30.342   65.000     60.515      47.279    2.996     11.304   364.687     49.209
  6.000     57.641   102.000   66.419      30.503   65.000     61.075      47.515    2.330     11.510   365.979     49.383
  6.000     57.641   102.000   67.647      30.669   65.000     61.647      47.758    1.664     11.722   367.324     49.564
  6.000     57.641   102.000   68.911      30.839   65.000     62.230      48.007    0.998     11.940   368.724     49.753
  6.000     57.641   102.000   70.214      31.015   65.000     62.824      48.263    0.332     12.165   370.180     49.950

AVERAGE COST OF GAS    48.470  CENTS PER MMBTU                 LEVELIZED COST OF GAS    48.088  CENTS PER MMBTU
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    ALGERIA GAS  AT  0.10/MMBTU   1-10-69    PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT       510.000          LIQUEFACTION PLANT CAPACITY 2320.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION     650.000           16    639.750  THOUSAND BARREL TANKERS
RECEIVING TERMINAL        88.800
TOTAL INVESTMENT        1248.800          GAS INPUT/OUTPUT RATIO       1.296

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT             OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS   OF GAS
CENTS PER   MILLIONS  COST      COSTS      CENTS PER   COST      COST    CENTS PER    COST       COST       OF     CENTS PER
 MMBTU     OF DOLLARS MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS    MMBTU

 10.000      96.070   102.000   51.000      33.608   65.000     53.316     49.573   12.986      8.845   389.218     52.519
 10.000      96.070   102.000   51.765      33.711   65.000     53.732     49.732   12.320      8.978   389.866     52.606
 10.000      96.070   102.000   52.552      33.817   65.000     54.156     49.896   11.654      9.814   390.549     52.696
 10.000      96.070   102.000   53.364      33.927   65.000     54.590     50.064   10.988      9.254   391.268     52.795
 10.000      96.070   102.000   54.200      34.040   65.000     55.031     50.236   10.322      9.398   392.024     52.897
 10.000      96.070   102.000   55.061      34.156   65.000     55.482     50.413    9.656      9.547   392.818     53.005
 10.000      96.070   102.000   55.948      34.276   65.000     55.942     50.595    8.990      9.700   393.652     53.117
 10.000      96.070   102.000   56.861      34.399   65.000     56.411     50.781    8.324      9.858   394.526     53.235
 10.000      96.070   102.000   57.802      34.525   65.000     56.889     50.973    7.658     10.021   395.442     53.354
 10.000      96.070   102.000   58.771      34.657   65.000     57.377     51.169    6.992     10.188   396.400     53.488
 10.000      96.070   102.000   59.769      34.791   65.000     57.874     51.371    6.326     10.361   397.402     53.623
 10.000      96.070   102.000   60.797      34.930   65.000     58.382     51.579    5.660     10.538   398.449     53.764
 10.000      96.070   102.000   61.856      35.073   65.000     58.899     51.791    4.994     10.721   399.543     53.912
 10.000      96.070   102.000   62.947      35.220   65.000     59.427     52.010    4.328     10.910   400.684     54.066
 10.000      96.070   102.000   64.070      35.372   65.000     59.966     52.234    3.662     11.104   401.874     54.227
 10.000      96.070   102.000   65.228      35.528   65.000     60.515     52.464    2.996     11.304   403.115     54.394
 10.000      96.070   102.000   66.419      35.689   65.000     61.075     52.701    2.330     11.510   404.406     54.568
 10.000      96.070   102.000   67.647      35.854   65.000     61.647     52.943    1.664     11.722   405.752     54.750
 10.000      96.070   102.000   68.911      36.025   65.000     62.230     53.193    0.998     11.940   407.152     54.939
 10.000      96.070   102.000   70.214      36.201   65.000     62.824     53.449    0.332     12.165   408.608     55.135

AVERAGE COST OF GAS    53.655  CENTS PER MMBTU                LEVELIZED COST OF GAS    53.273  CENTS PER MMBTU
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    LIBYAN GAS  AT  0.06/MMBTU    1-10-69   PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT
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 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      455.000          LIQUEFACTION PLANT CAPACITY 2080.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    640.000            60    635.000  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       89.100
TOTAL INVESTMENT       1184.100          GAS INPUT/OUTPUT RATIO       1.300
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PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY        741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS       LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED    VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
CENTS PER  MILLIONS   COST     COSTS     CENTS PER   COST      COST     CENTS PER   COST      COST       OF      CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS      MMBTU

  6.000     57.836    91.000   45.500      26.222   64.000      52.288     41.914   13.030      8.794   332.449     44.859
  6.000     57.836    91.000   46.182      26.315   64.000      52.693     42.061   12.362      8.924   333.000     44.933
  6.000     57.836    91.000   46.885      26.409   64.000      53.107     42.211   11.694      9.058   333.582     45.012
  6.000     57.836    91.000   47.609      26.507   64.000      53.529     42.366   11.026      9.196   334.198     45.095
  6.000     57.836    91.000   48.355      26.608   64.000      53.960     42.525   10.357      9.338   334.849     45.182
  6.000     57.836    91.000   49.123      26.711   64.000      54.399     42.688    9.689      9.485   335.534     45.275
  6.000     57.836    91.000   49.914      26.818   64.000      54.847     42.855    9.021      9.635   336.256     45.372
  6.000     57.836    91.000   50.729      26.928   64.000      55.304     43.026    8.353      9.791   337.015     45.475
  6.000     57.836    91.000   51.568      27.041   64.000      55.770     43.203    7.684      9.951   337.812     45.582
  6.000     57.836    91.000   52.433      27.158   64.000      56.246     43.383    7.016     10.116   338.649     45.695
  6.000     57.836    91.000   53.324      27.278   64.000      56.730     43.569    6.348     10.286   339.526     45.814
  6.000     57.836    91.000   54.241      27.402   64.000      57.225     43.760    5.680     10.461   340.445     45.938
  6.000     57.836    91.000   55.186      27.529   64.000      57.730     43.955    5.011     10.641   341.406     46.067
  6.000     57.836    91.000   56.159      27.661   64.000      58.244     44.156    4.343     10.826   342.411     46.203
  6.000     57.836    91.000   57.161      27.796   64.000      58.769     44.362    3.675     11.018   343.461     46.345
  6.000     57.836    91.000   58.193      27.935   64.000      59.304     44.573    3.007     11.215   344.557     46.492
  6.000     57.836    91.000   59.256      28.079   64.000      59.851     44.791    2.338     11.417   345.701     46.647
  6.000     57.836    91.000   60.352      28.227   64.000      60.408     45.014    1.670     11.626   346.894     46.808
  6.000     57.836    91.000   61.480      28.379   64.000      60.976     45.242    1.002     11.841   348.137     46.976
  6.000     57.836    91.000   62.642      28.535   64.000      61.555     45.477    0.334     12.063   349.432     47.150

AVERAGE COST OF GAS    45.846  CENTS PER MMBTU                 LEVELIZED COST OF GAS    45.512  CENTS PER MMBTU
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    LIBYAN GAS  AT  0.08/MMBTU    1-10-69   PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT       455.000          LIQUEFACTION PLANT CAPACITY 2080.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION     640.000            60    635.000  THOUSAND BARREL TANKERS
RECEIVING TERMINAL        89.100
TOTAL INVESTMENT        1184.100          GAS INPUT/OUTPUT RATIO       1.300

PROJECT LIFE  20  YEARS           GAS DELIVERED    2030.400 BILLION BTU/DAY         741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT              OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED    VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
CENTS PER   MILLIONS  COST     COSTS      CENTS PER  COST       COST    CENTS PER    COST       COST       OF     CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS DOLLARS     MMBTU

  8.000       77.116    91.000  45.500      28.824  64.000      52.288     44.515   13.030     8.794   351.728     47.460
  8.000       77.116    91.000  46.182      28.916  64.000      52.693     44.662   12.362     8.924   352.279     47.534
  8.000       77.116    91.000  46.885      29.011  64.000      53.107     44.813   11.694     9.058   352.861     47.613
  8.000       77.116    91.000  47.609      29.108  64.000      53.529     44.967   11.026     9.196   353.477     47.696
  8.000       77.116    91.000  48.355      29.209  64.000      53.960     45.126   10.357     9.338   354.128     47.784
  8.000       77.116    91.000  49.123      29.313  64.000      54.399     45.289    9.689     9.485   354.813     47.876
  8.000       77.116    91.000  49.914      29.420  64.000      54.847     45.456    9.021     9.635   355.535     47.974
  8.000       77.116    91.000  50.729      29.529  64.000      55.304     45.628    8.353     9.791   356.294     48.076
  8.000       77.116    91.000  51.568      29.643  64.000      55.770     45.804    7.684     9.951   357.091     48.184
  8.000       77.116    91.000  52.433      29.759  64.000      56.246     45.985    7.016    10.116   357.928     48.297
  8.000       77.116    91.000  53.324      29.880  64.000      56.730     46.170    6.348    10.286   358.805     48.415
  8.000       77.116    91.000  54.241      30.003  64.000      57.225     46.361    5.680    10.461   359.724     48.539
  8.000       77.116    91.000  55.186      30.131  64.000      57.730     46.557    5.011    10.641   360.685     48.669
  8.000       77.116    91.000  56.159      30.262  64.000      58.244     46.757    4.343    10.826   361.690     48.804
  8.000       77.116    91.000  57.161      30.397  64.000      58.769     46.963    3.675    11.018   362.740     48.946
  8.000       77.116    91.000  58.193      30.537  64.000      59.304     47.175    3.007    11.215   363.836     49.094
  8.000       77.116    91.000  59.256      30.680  64.000      59.851     47.392    2.338    11.417   364.980     49.248
  8.000       77.116    91.000  60.352      30.828  64.000      60.408     47.615    1.670    11.626   366.173     49.409
  8.000       77.116    91.000  61.480      30.980  64.000      60.976     47.844    1.002    11.841   367.416     49.577
  8.000       77.116    91.000  62.642      31.137  64.000      61.555     48.079    0.334    12.063   368.711     49.752

AVERAGE COST OF GAS    48.447  CENTS PER MMBTU                  LEVELIZED COST OF GAS    48.113  CENTS PER MMBTU
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8926

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    LIBYAN GAS  AT  0.10/MMBTU    1-1  69   PRO-. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      455.000          LIQUEFACTION PLANT CAPACITY 2080.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    640.000            60    635.000  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       89.100
TOTAL INVESTMENT       1184.100           GAS INPUT/OUTPUT RATIO        1.300

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY         741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED    VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST  FIXED   VARIABLE   MILLIONS   OF GAS
CENTS PER   MILLIONS  COST     COSTS     CENTS PER   COST      COST    CENTS PER   COST      COST        OF     CENTS PER
  MMBTU   OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU   MILLIONS OF DOLLARS DOLLARS    MMBTU

 10.000     96.394   91.000   45.500     31.425   64.000     52.288     47.117   13.030     8.794   371.007    50.062
 10.000     96.394   91.000   46.182     31.517   64.000     52.693     47.263   12.362     8.924   371.558    50.136
 10.000     96.394   91.000   46.885     31.612   64.000     53.107     47.414   11.694     9.058   372.140    50.214
 10.000     96.394   91.000   47.609     31.710   64.000     53.529     47.569   11.026     9.196   372.756    50.298
 10.000     96.394   91.000   48.355     31.811   64.000     53.960     47.728   10.357     9.338   373.407    50.385
 10.000     96.394   91.000   49.123     31.914   64.000     54.399     47.890    9.689     9.485   374.092    50.478
 10.000     96.394   91.000   49.914     32.021   64.000     54.847     48.058    9.021     9.635   374.814    50.575
 10.000     96.394   91.000   50.729     32.131   64.000     55.304     48.229    8.353     9.791   375.573    50.678
 10.000     96.394   91.000   51.568     32.244   64.000     55.770     48.405    7.684     9.951   376.370    50.785
 10.000     96.394   91.000   52.433     32.361   64.000     56.246     48.586    7.016    10.116   377.207    50.898
 10.000     96.394   91.000   53.324     32.481   64.000     56.730     48.772    6.348    10.286   378.084    51.016
 10.000     96.394   91.000   54.241     32.605   64.000     57.225     48.962    5.680    10.461   379.003    51.140
 10.000     96.394   91.000   55.186     32.732   64.000     57.730     49.158    5.011    10.641   379.964    51.270
 10.000     96.394   91.000   56.159     32.864   64.000     58.244     49.359    4.343    10.826   380.969    51.406
 10.000     96.394   91.000   57.161     32.999   64.000     58.769     49.565    3.675    11.018   382.019    51.547
 10.000     96.394   91.000   58.193     33.138   64.000     59.304     49.776    3.007    11.215   383.115    51.695
 10.000     96.394   91.000   59.256     33.282   64.000     59.851     49.993    2.338    11.417   384.259    51.850
 10.000     96.394   91.000   60.352     33.429   64.000     60.408     50.216    1.670    11.626   385.452    52.011
 10.000     96.394   91.000   61.480     33.582   64.000     60.976     50.445    1.002    11.841   386.695    52.178
 10.000     96.394   91.000   62.642     33.738   64.000     61.555     50.680    0.334    12.063   387.990    52.353

AVERAGE COST OF GAS    51.049  CENTS PER MMBTU                  LEVELIZED COST OF GAS    50.715  CENTS PER MMBTU
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     NIGERIAN GAS  AT  0.066/MMBTU  1-10-69  PROJ. 8926

   LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

   TANKER ESCALATION FACTOR                   2.000  PERCENT

   RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

    TOTAL PROJECT INVESTMENT
   LIQUEFACTION PLANT      615.000          LIQUEFACTION PLANT CAPACITY 2850.000 MILLION CUBIC FEET PER DAY
   OCEAN TRANSPORTATION    885.000            22     631.700  THOUSAND BARREL TANKERS
   RECEIVING TERMINAL       90.500
   TOTAL INVESTMENT       1590.500          GAS INPUT/OUTPUT RATIO        1.346

   PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

    TOTAL PURCHASED GAS       LIQUEFACTION PLANT              OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
   UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE-  MILLIONS    OF GAS
   CENTS PER   MILLIONS  COST      COSTS      CENTS PER  COST       COST    CENTS PER   COST       COST       OF      CENTS PER
    MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS      MMBTU

     6.600    65.868   123.000   61.500      33.783   88.500    72.377     55.491    13.235     8.998    433.478     58.491
     6.600    65.868   123.000   62.422      33.907   88.500    72.939     55.691    12.556     9.132    434.419     58.618
     6.600    65.868   123.000   63.372      34.036   88.500    73.513     55.897    11.878     9.270    435.402     58.751
     6.600    65.868   123.000   64.351      34.168   88.500    74.098     56.108    11.199     9.412    436.430     58.889
     6.600    65.868   123.000   65.359      34.304   88.500    74.695     56.325    10.520     9.559    437.502     59.034
     6.600    65.868   123.000   66.397      34.444   88.500    75.304     56.547     9.841     9.710    438.622     59.185
     6.600    65.868   123.000   67.467      34.588   88.500    75.925     56.775     9.163     9.865    439.789     59.343
     6.600    65.868   123.000   58.568      34.737   88.500    76.559     57.009     8.484    10.026    441.006     59.507
     6.600    65.868   123.000   69.703      34.890   88.500    77.205     57.249     7.805    10.191    442.273     59.678
     6.600    65.868   123.000   70.871      35.048   88.500    77.864     57.496     7.126    10.361    443.592     59.856
     6.600    65.868   123.000   72.075      35.210   88.500    78.536     57.749     6.448    10.536    444.964     60.041
     6.600    65.868   123.000   73.315      35.377   88.500    79.222     58.009     5.769    10.716    446.391     60.233
     6.600    65.868   123.000   74.592      35.550   88.500    79.921     58.276     5.090    10.902    447.874     60.434
     6.600    65.868   123.000   75.907      35.727   88.500    80.635     58.549     4.411    11.093    449.416     60.642
     6.600    65.868   123.000   77.262      35.910   88.500    81.362     58.830     3.733    11.290    451.016     60.858
     6.600    65.868   123.000   78.657      36.098   88.500    82.105     59.119     3.054    11.493    452.678     61.082
     6.600    65.868   123.000   80.094      36.292   88.500    82.862     59.415     2.375    11.702    454.403     61.315
     6.600    65.868   123.000   81.574      36.492   88.500    83.634     59.719     1.696    11.918    456.192     61.556
     6.600    65.868   123.000   83.099      36.698   88.500    84.422     60.031     1.018    12.139    458.047     61.806
     6.600    65.868   123.000   84.670      36.909   88.500    85.225     60.351     0.339    12.368    459.971     62.066

   AVERAGE COST OF GAS     60.069  CENTS PER MMBTU                LEVELIZED COST OF GAS    59.545  CENTS PER MMBTU
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     NIGERIAN GAS  AT  0.08 /MMBTU  1-10-69  PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      615.000          LIQUEFACTION PLANT CAPACITY 2850.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    885.000            22     631.700  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       90.500
TOTAL INVESTMENT       1590.500          GAS INPUT/OUTPUT RATIO        1.346

PROJECT LIFE  20  YEARS           GAS DELIVERED    2030.400 BILLION BTU/DAY           741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT               OCEAN TRANSPORTATION       RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED    VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
CENTS PER   MILLIONS  COST     COSTS     CENTS PER   COST       COST    CENTS PER   COST       COST       OF     CENTS PER
 MMBTU     OF DOLLARS MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS  MMBTU    MILLIONS OF DOLLARS DOLLARS     MMBTU

  8.000       79.839   123.000   61.500      35.668   88.500      72.377     57.376  13.235     8.998  447.450     60.376
  8.000       79.839   123.000   62.422      35.793   88.500      72.939     57.577  12.556     9.132  448.391     60.503
  8.000       79.839   123.000   63.372      35.921   88.500      73.513     57.782  11.878     9.270  449.374     60.636
  8.000       79.839   123.000   64.351      36.053   88.500      74.098     57.993  11.199     9.412  450.402     60.775
  8.000       79.839   123.000   65.359      36.189   88.500      74.695     58.210  10.520     9.559  451.474     60.919
  8.000       79.839   123.000   66.397      36.329   88.500      75.304     58.432   9.841     9.710  452.594     61.070
  8.000       79.839   123.000   67.467      36.473   88.500      75.925     58.660   9.163     9.865  453.761     61.228
  8.000       79.839   123.000   68.568      36.622   88.500      76.559     58.894   8.484    10.026  454.978     61.392
  8.000       79.839   123.000   69.703      36.775   88.500      77.205     59.135   7.805    10.191  456.245     61.563
  8.000       79.839   123.000   70.871      36.933   88.500      77.864     59.381   7.126    10.361  457.564     61.741
  8.000       79.839   123.000   72.075      37.095   88.500      78.536     59.634   6.448    10.536  458.936     61.926
  8.000       79.839   123.000   73.315      37.263   88.500      79.222     59.894   5.769    10.716  460.363     62.119
  8.000       79.839   123.000   74.592      37.435   88.500      79.921     60.161   5.090    10.902  461.846     62.319
  8.000       79.839   123.000   75.907      37.612   88.500      80.635     60.435   4.411    11.093  463.388     62.527
  8.000       79.839   123.000   77.262      37.795   88.500      81.362     60.716   3.733    11.290  464.988     62.743
  8.000       79.839   123.000   78.657      37.983   88.500      82.105     61.004   3.054    11.493  466.650     62.967
  8.000       79.839   123.000   80.094      38.177   88.500      82.862     61.300   2.375    11.702  468.375     63.200
  8.000       79.839   123.000   81.574      38.377   88.500      83.634     61.604   1.696    11.918  470.164     63.441
  8.000       79.839   123.000   83.099      38.583   88.500      84.422     61.916   1.018    12.139  472.019     63.692
  8.000       79.839   123.000   84.670      38.795   88.500      85.225     62.236   0.339    12.368  473.943     63.951

AVERAGE COST OF GAS    61.954  CENTS PER MMBTU                  LEVELIZED COST OF GAS    61.431  CENTS PER MMBTU
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    NIGERIAN GAS  AT  0.10 /MMBTU  1-10-69  PROJ. 8926

 LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

 TANKER ESCALATION FACTOR                   2.000  PERCENT

 RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

  TOTAL PROJECT INVESTMENT
 LIQUEFACTION PLANT      615.000        LIQUEFACTION PLANT CAPACITY 2850.000 MILLION CUBIC FEET PER DAY
 OCEAN TRANSPORTATION    885.000          22    631.700  THOUSAND BARREL TANKERS
 RECEIVING TERMINAL       90.500
 TOTAL INVESTMENT       1590.500        GAS INPUT/OUTPUT RATIO       1.346

 PROJECT LIFE  20  YEARS         GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

  TOTAL PURCHASED GAS       LIQUEFACTION PLANT           OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
 UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS   OF GAS
 CENTS PER   MILLIONS  COST      COSTS      CENTS PER  COST       COST    CENTS PER    COST      COST        OF     CENTS PER
  MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS     MMBTU

  10.000      99.800   123.000   61.500      38.362   88.500      72.377     60.070   13.235     8.998   467.410     63.079
  10.000      99.800   123.000   62.422      38.486   88.500      72.939     60.270   12.556     9.132   468.351     63.197
  10.000      99.800   123.000   63.372      38.614   88.500      73.513     60.476   11.878     9.270   469.334     63.329
  10.000      99.800   123.000   64.351      38.746   88.500      74.098     60.687   11.199     9.412   470.362     63.468
  10.000      99.800   123.000   65.359      38.882   88.500      74.695     60.903   10.520     9.559   471.434     63.613
  10.000      99.800   123.000   66.397      39.022   88.500      75.304     61.125    9.841     9.710   472.554     63.764
  10.000      99.800   123.000   67.467      39.167   88.500      75.925     61.354    9.163     9.865   473.721     63.921
  10.000      99.800   123.000   68.568      39.315   88.500      76.559     61.588    8.484    10.026   474.938     64.085
  10.000      99.800   123.000   69.703      39.468   88.500      77.205     61.828    7.805    10.191   476.205     64.256
  10.000      99.800   123.000   70.871      39.626   88.500      77.864     62.075    7.126    10.361   477.524     64.434
  10.000      99.800   123.000   72.075      39.789   88.500      78.536     62.328    6.448    10.536   478.896     64.620
  10.000      99.800   123.000   73.315      39.956   88.500      79.222     62.588    5.769    10.716   480.323     64.812
  10.000      99.800   123.000   74.592      40.128   88.500      79.921     62.854    5.090    10.902   481.806     65.012
  10.000      99.800   123.000   75.907      40.306   88.500      80.635     63.128    4.411    11.093   483.348     65.220
  10.000      99.800   123.000   77.262      40.488   88.500      81.362     63.409    3.733    11.290   484.948     65.436
  10.000      99.800   123.000   78.657      40.677   88.500      82.105     63.697    3.054    11.493   486.610     65.660
  10.000      99.800   123.000   80.094      40.871   88.500      82.862     63.993    2.375    11.702   488.335     65.893
  10.000      99.800   123.000   81.574      41.070   88.500      83.634     64.297    1.696    11.918   490.124     66.135
  10.000      99.800   123.000   83.099      41.276   88.500      84.422     64.609    1.018    12.139   491.979     66.385
  10.000      99.800   123.000   84.670      41.488   88.500      85.225     64.930    0.339    12.368   493.903     66.645

 AVERAGE COST OF GAS    64.648  CENTS PER MMBTU              LEVELIZED COST OF GAS    64.124  CENTS PER MMBTU
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8926

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    CANADIAN GAS  AT  0.12/MMBTU  1-10-69   PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      625.000        LIQUEFACTION PLANT CAPACITY 3000.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    385.000            9    608.000  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       98.900
TOTAL INVESTMENT       1108.900        GAS INPUT/OUTPUT RATIO       1.239

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT              OCEAN TRANSPORTATION       RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED     VARIABLE   MILLIONS   OF GAS
CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST       COST    CENTS PER    COST       COST        OF      CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS      MMBTU

 12.000     110.195   125.000   62.500      40.169   38.500     32.459     49.744   14.464     9.739   392.858     53.010
 12.000     110.195   125.000   63.437      40.296   38.500     32.723     49.906   13.722     9.882   393.461     53.091
 12.000     110.195   125.000   64.403      40.426   38.500     32.992     50.073   12.980    10.030   394.103     53.178
 12.000     110.195   125.000   65.397      40.560   38.500     33.267     50.244   12.238    10.183   394.783     53.270
 12.000     110.195   125.000   66.422      40.698   38.500     33.547     50.420   11.497    10.340   395.503     53.367
 12.000     110.195   125.000   67.477      40.841   38.500     33.833     50.601   10.755    10.502   396.265     53.470
 12.000     110.195   125.000   68.564      40.987   38.500     34.125     50.787   10.013    10.669   397.068     53.578
 12.000     110.195   125.000   69.683      41.139   38.500     34.422     50.978    9.271    10.841   397.915     53.692
 12.000     110.195   125.000   70.836      41.294   38.500     34.726     51.175    8.530    11.017   398.806     53.813
 12.000     110.195   125.000   72.024      41.454   38.500     35.035     51.377    7.788    11.200   399.744     53.939
 12.000     110.195   125.000   73.247      41.619   38.500     35.351     51.585    7.046    11.387   400.729     54.072
 12.000     110.195   125.000   74.507      41.789   38.500     35.673     51.798    6.304    11.581   401.762     54.211
 12.000     110.195   125.000   75.804      41.964   38.500     36.002     52.017    5.563    11.780   402.846     54.358
 12.000     110.195   125.000   77.141      42.145   38.500     36.337     52.243    4.821    11.985   403.981     54.511
 12.000     110.195   125.000   78.518      42.331   38.500     36.678     52.475    4.079    12.196   405.169     54.671
 12.000     110.195   125.000   79.936      42.522   38.500     37.027     52.713    3.337    12.413   406.411     54.839
 12.000     110.195   125.000   81.396      42.719   38.500     37.383     52.958    2.596    12.637   407.710     55.014
 12.000     110.195   125.000   82.901      42.922   38.500     37.745     53.210    1.854    12.868   409.066     55.197
 12.000     110.195   125.000   84.450      43.131   38.500     38.115     53.469    1.112    13.106   410.481     55.388
 12.000     110.195   125.000   86.046      43.347   38.500     38.492     53.736    0.370    13.351   411.957     55.587

AVERAGE COST OF GAS    54.113  CENTS PER MMBTU              LEVELIZED COST OF GAS    53.738  CENTS PER MMBTU
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    CANADIAN GAS  AT  0.10/MMBTU  1-10-69   PROJ. 8926

 LIQUEFACTION COST ESCALATION FACTOR       3.000  PERCENT

 TANKER ESCALATION FACTOR                  2.000  PERCENT

 RECEIVING TERMINAL ESCALATION FACTOR      3.000  PERCENT

  TOTAL PROJECT INVESTMENT
 LIQUEFACTION PLANT      625.000         LIQUEFACTION PLANT CAPACITY 3000.000 MILLION CUBIC FEET PER DAY
 OCEAN TRANSPORTATION    385.000            9    608.000  THOUSAND BARREL TANKERS
 RECEIVING TERMINAL       98.900
 TOTAL INVESTMENT       1108.900         GAS INPUT/OUTPUT RATIO       1.239

 PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

  TOTAL PURCHASED GAS       LIQUEFACTION PLANT          OCEAN TRANSPORTATION     RECEIVING TERMINAL TOTAL COST UNIT COST
 UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
 CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST      COST    CENTS PER    COST       COST       OF     CENTS PER
  MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU   MILLIONS OF DOLLARS DOLLARS     MMBTU

  10.000      91.830   125.000    62.500      37.691   38.500      32.459    47.266   14.464      9.739   374.492     50.532
  10.000      91.830   125.000    63.437      37.817   38.500      32.723    47.428   13.722      9.882   375.095     50.613
  10.000      91.830   125.000    64.403      37.948   38.500      32.992    47.595   12.980     10.030   375.737     50.700
  10.000      91.830   125.000    65.397      38.082   38.500      33.267    47.766   12.238     10.183   376.417     50.792
  10.000      91.830   125.000    66.422      38.220   38.500      33.547    47.942   11.497     10.340   377.137     50.889
  10.000      91.830   125.000    67.477      38.363   38.500      33.833    48.123   10.755     10.502   377.898     50.991
  10.000      91.830   125.000    68.564      38.509   38.500      34.125    48.309   10.013     10.669   378.702     51.100
  10.000      91.830   125.000    69.683      38.660   38.500      34.422    48.500    9.271     10.841   379.549     51.214
  10.000      91.830   125.000    70.836      38.816   38.500      34.726    48.697    8.530     11.017   380.440     51.334
  10.000      91.830   125.000    72.024      38.976   38.500      35.035    48.899    7.788     11.200   381.378     51.461
  10.000      91.830   125.000    73.247      39.141   38.500      35.351    49.106    7.046     11.387   382.363     51.594
  10.000      91.830   125.000    74.507      39.311   38.500      35.673    49.320    6.304     11.581   383.396     51.733
  10.000      91.830   125.000    75.804      39.486   38.500      36.002    49.539    5.563     11.780   384.480     51.879
  10.000      91.830   125.000    77.141      39.667   38.500      36.337    49.765    4.821     11.985   385.615     52.033
  10.000      91.830   125.000    78.516      39.852   38.500      36.678    49.997    4.079     12.196   386.803     52.193
  10.000      91.830   125.000    79.936      40.044   38.500      37.027    50.235    3.337     12.413   388.045     52.361
  10.000      91.830   125.000    81.396      40.241   38.500      37.383    50.480    2.596     12.637   389.344     52.536
  10.000      91.830   125.000    82.901      40.444   38.500      37.745    50.732    1.854     12.868   390.700     52.719
  10.000      91.830   125.000    84.450      40.653   38.500      38.115    50.991    1.112     13.106   392.115     52.910
  10.000      91.830   125.000    86.046      40.868   38.500      38.492    51.257    0.370     13.351   393.592     53.109

 AVERAGE COST OF GAS    51.635  CENTS PER MMBTU                LEVELIZED COST OF GAS    51.260  CENTS PER MMBTU
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     CANADIAN GAS  AT  0.08/MMBTU  1-10-69   PROJ. 8926

   LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

   TANKER ESCALATION FACTOR                   2.000  PERCENT

   RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

    TOTAL PROJECT INVESTMENT
   LIQUEFACTION PLANT      625.000          LIQUEFACTION PLANT CAPACITY 3000.000 MILLION CUBIC FEET PER DAY
   OCEAN TRANSPORTATION    385.000             9     608.000  THOUSAND BARREL TANKERS
   RECEIVING TERMINAL       98.900
   TOTAL INVESTMENT       1108.900          GAS INPUT/OUTPUT RATIO       1.239

   PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

    TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION       RECEIVING TERMINAL TOTAL COST UNIT COST
   UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
   CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST      COST    CENTS PER    COST      COST        OF      CENTS PER
    MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU     MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS     MMBTU

     8.000      73.464   125.000  62.500      35.213   38.500     32.459     44.788   14.464      9.739  356.126     48.054
     8.000      73.464   125.000  63.437      35.339   38.500     32.723     44.950   13.722      9.882  356.729     48.135
     8.000      73.464   125.000  64.403      35.470   38.500     32.992     45.116   12.980     10.030  357.371     48.222
     8.000      73.464   125.000  65.397      35.604   38.500     33.267     45.288   12.238     10.183  358.051     48.313
     8.000      73.464   125.000  66.422      35.742   38.500     33.547     45.464   11.497     10.340  358.771     48.410
     8.000      73.464   125.000  67.477      35.884   38.500     33.833     45.645   10.755     10.502  359.533     48.513
     8.000      73.464   125.000  68.564      36.031   38.500     34.125     45.831   10.013     10.669  360.336     48.622
     8.000      73.464   125.000  69.683      36.182   38.500     34.422     46.022    9.271     10.841  361.183     48.736
     8.000      73.464   125.000  70.836      36.338   38.500     34.726     46.218    8.530     11.017  362.074     48.856
     8.000      73.464   125.000  72.024      36.498   38.500     35.035     46.420    7.788     11.200  363.012     48.983
     8.000      73.464   125.000  73.247      36.663   38.500     35.351     46.628    7.046     11.387  363.997     49.116
     8.000      73.464   125.000  74.507      36.833   38.500     35.673     46.842    6.304     11.581  365.030     49.255
     8.000      73.464   125.000  75.804      37.008   38.500     36.002     47.061    5.563     11.780  366.114     49.401
     8.000      73.464   125.000  77.141      37.188   38.500     36.337     47.287    4.821     11.985  367.249     49.554
     8.000      73.464   125.000  78.518      37.374   38.500     36.678     47.518    4.079     12.196  368.437     49.715
     8.000      73.464   125.000  79.936      37.566   38.500     37.027     47.757    3.337     12.413  369.679     49.882
     8.000      73.464   125.000  81.396      37.763   38.500     37.383     48.002    2.596     12.637  370.978     50.058
     8.000      73.464   125.000  82.901      37.966   38.500     37.745     48.254    1.854     12.868  372.334     50.241
     8.000      73.464   125.000  84.450      38.175   38.500     38.115     48.513    1.112     13.106  373.749     50.431
     8.000      73.464   125.000  86.046      38.390   38.500     38.492     48.779    0.370     13.351  375.226     50.631

   AVERAGE COST OF GAS    49.156   CENTS PER MMBTU              LEVELIZED COST OF GAS    48.781  CENTS PER MMBTU
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      CANADIAN GAS  AT  0.06/MMBTU  1-10-69   PROJ. 8926

   LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

   TANKER ESCALATION FACTOR                   2.000  PERCENT

   RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

    TOTAL PROJECT INVESTMENT
   LIQUEFACTION PLANT      625.000          LIQUEFACTION PLANT CAPACITY 3000.000 MILLION CUBIC FEET PER DAY
   OCEAN TRANSPORTATION    385.000              9    608.000  THOUSAND BARREL TANKERS
   RECEIVING TERMINAL       98.900
   TOTAL INVESTMENT       1108.900          GAS INPUT/OUTPUT RATIO        1.239

   PROJECT LIFE  20  YEARS           GAS DELIVERED    2030.400 MILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

    TOTAL PURCHASED GAS        LIQUEFACTION PLANT                 OCEAN TRANSPORTATION       RECEIVING TERMINAL TOTAL COST UNIT COST
   UNIT COST TOTAL COST  FIXED      VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED     VARIABLE  MILLIONS   OF GAS
   CENTS PER   MILLIONS  COST       COSTS     CENTS PER   COST      COST   CENTS PER     COST       COST       OF     CENTS PER
    MMBTU   OF DOLLARS MILLIONS OF DOLLARS   MMBTU     MILLIONS OF DOLLARS   MMBTU     MILLIONS OF DOLLARS  DOLLARS    MMBTU

    6.000     55.097   125.000   62.500     32.735    38.500    32.459    42.309    14.464     9.739   337.759   45.576
    6.000     55.097   125.000   63.437     32.861    38.500    32.723    42.472    13.722     9.882   338.361   45.657
    6.000     55.097   125.000   64.403     32.991    38.500    32.992    42.638    12.980    10.030   339.002   45.743
    6.000     55.097   125.000   65.397     33.126    38.500    33.267    42.810    12.238    10.183   339.682   45.835
    6.000     55.097   125.000   66.422     33.264    38.500    33.547    42.986    11.497    10.340   340.403   45.932
    6.000     55.097   125.000   67.477     33.406    38.500    33.833    43.167    10.755    10.502   341.167   46.035
    6.000     55.097   125.000   68.564     33.553    38.500    34.125    43.353    10.013    10.669   341.970   46.144
    6.000     55.097   125.000   69.683     33.704    38.500    34.422    43.544     9.271    10.841   342.817   46.258
    6.000     55.097   125.000   70.836     33.859    38.500    34.726    43.740     8.530    11.017   343.715   46.378
    6.000     55.097   125.000   72.024     34.020    38.500    35.035    43.942     7.788    11.200   344.644   46.505
    6.000     55.097   125.000   73.247     34.185    38.500    35.351    44.150     7.046    11.387   345.631   46.637
    6.000     55.097   125.000   74.507     34.355    38.500    35.673    44.363     6.304    11.581   346.664   46.777
    6.000     55.097   125.000   75.804     34.530    38.500    36.002    44.583     5.563    11.780   347.748   46.923
    6.000     55.097   125.000   77.141     34.710    38.500    36.337    44.808     4.821    11.985   348.883   47.076
    6.000     55.097   125.000   78.510     34.896    38.500    36.678    45.040     4.079    12.196   350.071   47.237
    6.000     55.097   125.000   79.936     35.087    38.500    37.027    45.279     3.337    12.413   351.413   47.418
    6.000     55.097   125.000   81.396     35.284    38.500    37.383    45.524     2.596    12.637   352.617   47.579
    6.000     55.097   125.000   82.901     35.487    38.500    37.745    45.776     1.854    12.868   353.968   47.763
    6.000     55.097   125.000   84.450     35.696    38.500    38.115    46.035     1.112    13.106   355.383   47.954
    6.000     55.097   125.000   86.046     35.912    38.500    38.492    46.301     0.370    13.351   356.856   48.153

    AVERAGE COST OF GAS     46.676  CENTS PER MMBTU               LEVELIZED COST OF GAS    46.303  CENTS PER MMBTU
    ALASKAN GAS  AT   0.15/MMBTU   1-10-69   PROJ. 8926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      615.000           LIQUEFACTION PLANT CAPACITY 2860.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    450.000             10    643.650  THOUSAND BARREL TANKERS
RECEIVING TERMINAL       94.400
TOTAL INVESTMENT       1159.400           GAS INPUT/OUTPUT RATIO       1.250

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL  TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS   OF GAS
CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST      COST    CENTS PER    COST       COST     . OF     CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS   MMBTU     MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS    MMBTU

  14.999     138.974   123.000   61.500      43.648   45.000      36.650     54.665  13.805      9.340   428.270     57.788
  14.999     138.974   123.000   62.422      43.772   45.000      36.933     54.828  13.097      9.478   428.906     57.874
  14.999     138.974   123.000   63.372      43.900   45.000      37.221     54.995  12.389      9.621   429.580     57.965
  14.999     138.974   123.000   64.351      44.032   45.000      37.516     55.167  11.681      9.768   430.292     58.061
  14.999     138.974   123.000   65.359      44.160   45.000      37.816     55.343  10.973      9.919   431.044     58.163
  14.999     138.974   123.000   66.397      44.309   45.000      38.122     55.525  10.265     10.075   431.837     58.270
  14.999     138.974   123.000   67.467      44.453   45.000      38.435     55.711   9.557     10.236   432.671     58.382
  14.999     138.974   123.000   68.568      44.601   45.000      38.753     55.903   8.849     10.402   433.549     58.501
  14.999     138.974   123.000   69.703      44.755   45.000      39.078     56.100   8.141     10.572   434.471     58.625
  14.999     138.974   123.000   70.871      44.912   45.000      39.410     56.302   7.433     10.748   435.439     58.756
  14.999     138.974   123.000   72.075      45.075   45.000      39.748     56.510   6.725     10.928   436.453     58.893
  14.999     138.974   123.000   73.315      45.242   45.000      40.093     56.724   6.017     11.115   437.516     59.036
  14.999     138.974   123.000   74.592      45.414   45.000      40.445     56.944   5.309     11.307   438.629     59.186
  14.999     138.974   123.000   75.907      45.592   45.000      40.804     57.170   4.601     11.504   439.793     59.343
  14.999     138.974   123.000   77.262      45.775   45.000      41.170     57.402   3.893     11.708   441.009     59.507
  14.999     138.974   123.000   78.657      45.963   45.000      41.544     57.641   3.185     11.917   442.280     59.679
  14.999     138.974   123.000   80.094      46.157   45.000      41.924     57.886   2.477     12.133   443.606     59.858
  14.999     138.974   123.000   81.574      46.357   45.000      42.313     58.138   1.769     12.356   444.989     60.044
  14.999     138.974   123.000   83.099      46.562   45.000      42.709     58.397   1.061     12.585   446.431     60.239
  14.999     138.974   123.000   84.670      46.774   45.000      43.113     58.564   0.353     12.821   447.934     60.442

AVERAGE COST OF GAS     58.930  CENTS PER MMBTU                LEVELIZED COST OF GAS    58.544  CENTS PER MMBTU
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    ALASKAN GAS  AT   0.10/MMBTU   1-10-69   PROJ. 6926

LIQUEFACTION COST ESCALATION FACTOR        3.000  PERCENT

TANKER ESCALATION FACTOR                   2.000  PERCENT

RECEIVING TERMINAL ESCALATION FACTOR       3.000  PERCENT

 TOTAL PROJECT INVESTMENT
LIQUEFACTION PLANT      615.000          LIQUEFACTION PLANT CAPACITY 2860.000 MILLION CUBIC FEET PER DAY
OCEAN TRANSPORTATION    450.000            10    643.650  THOUSAND BARREL TANKERS              ,
RECEIVING TERMINAL       94.400
TOTAL INVESTMENT       1159.400          GAS INPUT/OUTPUT RATIO        1.250

PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

 TOTAL PURCHASED GAS        LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED   VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
CENTS PER   MILLIONS  COST       COSTS    CENTS PER   COST      COST    CENTS PER    COST       COST       OF     CENTS PER
 MMBTU    OF DOLLARS MILLIONS OF DOLLARS  MMBTU     MILLIONS OF DOLLARS  MMBTU    MILLIONS OF DOLLARS DOLLARS      MMBTU

 10.000     92.650   123.000   61.500      37.397   45.000     36.650    48.414   13.805     9.340   381.945     51.547
 10.000     92.650   123.000   62.422      37.521   45.000     36.933    48.577   13.097     9.478   382.582     51.625
 10.000     92.650   123.000   63.372      37.650   45.000     37.221    48.744   12.389     9.621   383.255     51.714
 10.000     92.650   123.000   64.351      37.782   45.000     37.516    48.916   11.681     9.768   383.967     51.810
 10.000     92.650   123.000   65.359      37.918   45.000     37.816    49.042   10.973     9.919   384.719     51.912
 10.000     92.650   123.000   66.397      38.058   45.000     38.122    49.274   10.265    10.075   385.512     52.019
 10.000     92.650   123.000   67.467      38.202   45.000     38.435    49.460    9.557    10.236   386.346     52.131
 10.000     92.650   123.000   68.568      38.351   45.000     38.753    49.652    8.849    10.402   387.224     52.250
 10.000     92.650   123.000   69.703      38.504   45.000     39.078    49.849    8.141    10.572   388.146     52.374
 10.000     92.650   123.000   70.871      38.661   45.000     39.410    50.051    7.433    10.748   389.114     52.505
 10.000     92.650   123.000   72.075      38.824   45.000     39.748    50.259    6.725    10.928   390.128     52.642
 10.000     92.650   123.000   73.315      38.991   45.000     40.093    50.473    6.017    11.115   391.191     52.785
 10.000     92.650   123.000   74.592      39.163   45.000     40.445    50.693    5.309    11.307   392.304     52.935
 10.000     92.650   123.000   75.907      39.341   45.000     40.804    50.919    4.601    11.504   393.468     53.092
 10.000     92.650   123.000   77.262      39.524   45.000     41.170    51.151    3.893    11.708   394.684     53.256
 10.000     92.650   123.000   78.657      39.712   45.000     41.544    51.390    3.185    11.917   395.955     53.428
 10.000     92.650   123.000   80.094      39.906   45.000     41.924    51.635    2.477    12.133   397.281     53.607
 10.000     92.650   123.000   81.574      40.106   45.000     42.313    51.887    1.769    12.356   398.664     53.795
 10.000     92.650   123.000   83.099      40.311   45.000     42.709    52.146    1.061    12.585   400.106     53.988
 10.000     92.650   123.000   84.670      40.523   45.000     43.113    52.413    0.353    12.821   401.609     54.191

AVERAGE COST OF GAS    52.680  CENTS PER MMBTU                  LEVELIZED COST OF GAS    52.293  CENTS PER MMBTU
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   ALASKAN GAS  AT  0.15/MMBTU  US CREWS  1-10-69  PROJ.  8926

 LIQUEFACTION COST ESCALATION FACTOR       3.000  PERCENT

 TANKER ESCALATION FACTOR                  2.000  PERCENT

 RECEIVING TERMINAL ESCALATION FACTOR      3.000  PERCENT

  TOTAL PROJECT INVESTMENT
 LIQUEFACTION PLANT     615.000        LIQUEFACTION PLANT CAPACITY 2860.000 MILLION CUBIC FEET PER DAY
 OCEAN TRANSPORTATION   630.000          10    643.650  THOUSAND BARREL TANKERS
 RECEIVING TERMINAL      94.400
 TOTAL INVESTMENT      1339.400        GAS INPUT/OUTPUT RATIO       1.250

 PROJECT LIFE  20  YEARS          GAS DELIVERED    2030.400 BILLION BTU/DAY          741.096  MILLION MMBTU/YEAR

  TOTAL PURCHASED GAS       LIQUEFACTION PLANT            OCEAN TRANSPORTATION      RECEIVING TERMINAL TOTAL COST UNIT COST
 UNIT COST TOTAL COST  FIXED     VARIABLE F.O.B. COST  FIXED    VARIABLE C.I.F. COST   FIXED    VARIABLE  MILLIONS    OF GAS
 CENTS PER   MILLIONS  COST      COSTS     CENTS PER   COST       COST    CENTS PER    COST      COST       OF     CENTS PER
   MMBTU    OF DOLLARS MILLIONS  OF DOLLARS   MMBTU    MILLIONS OF DOLLARS   MMBTU    MILLIONS OF DOLLARS DOLLARS    MMBTU

  14.999     138.974   123.000   61.500      43.648   63.000     53.610     59.382   13.805     9.340   463.230    62.506
  14.999     138.974   123.000   62.422      43.772   63.000     54.052     59.567   13.097     9.478   464.026    62.613
  14.999     138.974   123.000   63.372      43.900   63.000     54.503     59.756   12.389     9.621   464.862    62.726
  14.999     138.974   123.000   64.351      44.032   63.000     54.963     59.950   11.681     9.768   465.740    62.844
  14.999     138.974   123.000   65.359      44.168   63.000     55.432     60.149   10.973     9.919   466.660    62.968
  14.999     138.974   123.000   66.397      44.309   63.000     55.911     60.354   10.265    10.075   467.625    63.099
  14.999     138.974   123.000   67.467      44.453   63.000     56.399     60.564    9.557    10.236   468.635    63.235
  14.999     138.974   123.000   68.568      44.601   63.000     56.897     60.780    8.849    10.402   469.692    63.378
  14.999     138.974   123.000   69.703      44.755   63.000     57.405     61.002    8.141    10.572   470.797    63.527
  14.999     138.974   123.000   70.871      44.912   63.000     57.923     61.229    7.433    10.748   471.952    63.682
  14.999     138.974   123.000   72.075      45.075   63.000     58.451     61.463    6.725    10.928   473.157    63.845
  14.999     138.974   123.000   73.315      45.242   63.000     58.990     61.703    6.017    11.115   474.414    64.015
  14.999     138.974   123.000   74.592      45.414   63.000     59.540     61.949    5.309    11.307   475.724    64.192
  14.999     138.974   123.000   75.907      45.592   63.000     60.101     62.202    4.601    11.504   477.090    64.376
  14.999     138.974   123.000   77.262      45.775   63.000     60.673     62.462    3.893    11.708   478.512    64.568
  14.999     138.974   123.000   78.657      45.963   63.000     61.257     62.729    3.185    11.917   479.993    64.768
  14.999     138.974   123.000   80.094      46.157   63.000     61.852     63.004    2.477    12.133   481.533    64.975
  14.999     138.974   123.000   81.574      46.357   63.000     62.459     63.285    1.769    12.356   483.135    65.191
  14.999     138.974   123.000   83.099      46.562   63.000     63.078     63.575    1.061    12.585   484.800    65.416
  14.999     138.974   123.000   84.670      46.774   63.000     63.710     63.872    0.353    12.821   486.530    65.650

 AVERAGE COST OF GAS    63.879  CENTS PER MMBTU                 LEVELIZED COST OF GAS    63.419  CENTS PER MMBTU
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Reference Cited

1. Texas Eastern Transmission Corp., Engineering Dept., Process Engineering Div., LNG Conversion Tables. Shreveport, La., 1968.